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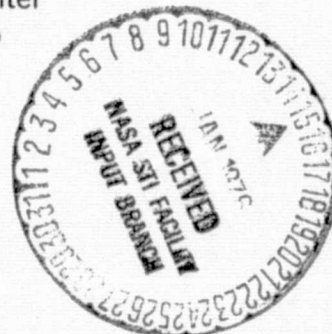
STUDIES OF THE USE OF HIGH-TEMPERATURE NUCLEAR HEAT FROM AN HTGR FOR HYDROGEN PRODUCTION

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L. J. Halvers, and A. M. Jahromi

GENERAL ATOMIC COMPANY

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1. ABSTRACT

This report presents the results of a study performed for the National Aeronautics and Space Administration which surveyed various methods of hydrogen production using nuclear and fossil energy. A description of these methods is provided, and efficiencies are calculated for each case.

The process designs of systems that utilize the heat from a General Atomic high-temperature gas-cooled reactor with a steam methane reformer, and feed the reformer with substitute natural gas manufactured from coal, using reforming temperatures of 922°K (1200°F), 1033°K (1400°F), 1144°K (1600°F), 1256°K (1800°F), and 1367°K (2000°F) are presented. The capital costs for these systems and the resultant hydrogen production price for these cases are discussed. Hydrogen production costs are shown to decrease rapidly with increasing reformer temperatures of 922°K (1200°F) to about 978°K (1500°F); costs show a less dramatic decrease from 978°K (1500°F) to 1367°K (2000°F). The hydrogen price is not sensitive to the cost of nuclear fuel but is sensitive to the cost of coal. It is concluded that the best process for development with a high probability of success is one which utilizes a reforming temperature of 1033°K (1400°F) although the "ideal" temperature may be between 1033°K (1400°F) and 1144°K (1600°F). A research and development program for commercialization is described.

2. INTRODUCTION AND SUMMARY

At the request of the National Aeronautics and Space Administration (NASA), a study was performed to assess the technology of several hydrogen production processes. In addition to a general survey, the production of hydrogen by hydrogasification of coal liquids using a high-temperature gas-cooled nuclear reactor (HTGR) was studied in detail. A market survey of the possible future demand for hydrogen is also included.

The following hydrogen processes are included in this report:

<u>Process</u>	<u>Feedstock</u>	<u>Heat Source</u>
Steam hydrocarbon reforming	Natural gas, water	Natural gas/nuclear
Hydrogasification of coal liquids	Coal, water	Nuclear
Lurgi gasification of coal	Coal, water	Coal
Steam coal gasification	Coal, water	Nuclear
Electrolysis of water	Water	Nuclear
Ispra Mark 10 water splitting	Water	Nuclear

Although several other processes are potential candidates for the future manufacture of hydrogen, the processes surveyed are considered to be a representative sample. They cover a technology range of current state-of-the-art hydrogen manufacturing processes to processes currently at primitive stages of development. Feedstocks include light hydrocarbons, coal, and water. Because hydrogen processes consume large quantities of energy, particular attention has been given to energy sources utilizing fossil and nuclear fuels.

Over the past 25 years, hydrogen has primarily been produced by steam hydrocarbon reforming of natural gas. However, the increasing scarcity of natural gas makes this a questionable source of future hydrogen. Lurgi

gasification of coal and electrolysis of water utilize more abundant natural resources with proven technology. These processes show substantial penalties in thermal efficiency and in the past have not been economically competitive with the reforming process.

Substitution of nuclear heat for conventional fossil-fired heaters in the steam hydrocarbon reforming process offers a substantial reduction of the natural gas required. Reforming conditions are compatible with the present capabilities of the HTGR, and a limited development effort would be required to achieve commercialization. The process can be coupled with hydrogasification of coal liquids (Stone & Webster/Gulf process) to produce hydrogen from coal.

The direct steam coal gasification process offers an efficient means of producing hydrogen from coal. The process temperatures required are over 1200°K (1700°F), and present nuclear technology can deliver only a limited portion of the reactor thermal power at these temperatures. A substantial development program concentrating on high-temperature materials is required to determine the feasibility of the process.

The ultimate future source of hydrogen will undoubtedly be closed cycle disassociation of water. The only proven process for production of hydrogen from water is the electrolytic process. A major drawback of the electrolytic process is its dependence on the low efficiencies associated with thermoelectric energy conversion. Conceivably, production of electric power with an efficiency of 50% and development of more efficient electrolytic cells could increase the process efficiency from 32% to as high as 49%.

Thermochemical water splitting offers the possibility of more efficient conversion of water to hydrogen than electrolysis. Any evaluation of water splitting cycles is difficult at their present stage of development. Several cycles are potential candidates, each opening a new

area of process technology. The Ispra Mark 10 cycle was selected for study in this report. Although the results of this study indicate that the thermal efficiency of this cycle is low, it is believed that other cycles could yield efficiencies which are considerably higher. It is clear that high separation efficiencies are crucial to achieving high thermal utilization. An extensive materials development effort will be required to achieve commercialization since high temperatures are needed in the process.

A comparison of the processes would certainly be difficult and is beyond the scope of this work. The processes cover a wide range of development status and a wide range of availability of feedstock and fuel. It is likely that several processes will reach commercialization in an appropriate sequence of time as fossil fuel resources become increasingly more scarce and costly and the technology for utilizing more abundant resources develops.

Gasification of coal liquids using a nuclear heat source was studied in more detail to determine the effect of process temperature (maximum reforming temperature) on the development status, efficiency, and economics of the process. The nuclear heat sources for this work were taken from a previous study (Ref. 2-1). The results of this study showed that the process efficiency increases and the hydrogen cost decreases over the entire temperature range studied, i.e., from 922°K (1200°F) to 1367°K (2000°F). Process efficiency and hydrogen cost are most sensitive to the temperature range 922° to 1089°K (1200° to 1500°F).

The development status of the nuclear heat source is strongly affected by process temperature. At process temperatures of 1033°K (1400°F) and below, a limited development effort is required. At process temperatures of 1144°K (1600°F) and above, a major development program in high-temperature materials is required. The development status of other process units is not affected by process temperature.

It is recommended that future work concentrate on a process temperature of about 1033°K (1400°F). At this temperature, with a nuclear fuel cost of \$0.44/GJ (\$/MMBtu), an Oklahoma coal cost of \$11/Kkg (\$10/ton), and a fixed charge rate of 15%, the price of hydrogen would be \$0.42/kg (\$1.07 mscf)* (July 1974 basis). Hydrogen price is insensitive to nuclear fuel cost but coal costs have a small but significant effect on hydrogen price.

A survey of future hydrogen markets indicates that the demand for hydrogen will significantly increase by the year 2000, particularly in the manufacture of synthetic fuels. Although it would appear that most of the hydrogen will be captively produced by the consuming plant, this does not necessarily exclude the development of a merchant hydrogen market.

REFERENCE

- 2-1. "High-Temperature Nuclear Heat Source Study," USAEC Report GA-A13158, General Atomic, December 30, 1974.

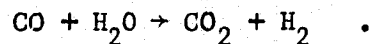
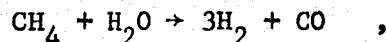
*At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

3. TECHNOLOGICAL ASSESSMENT OF SELECTED HYDROGEN PROCESSES

3.1. HYDROGEN FROM NATURAL GAS (PROCESS 1)

Natural gas is currently the most widely used feedstock for production of hydrogen in the United States. Two commercially acceptable processes exist: (1) steam-methane reforming and (2) partial oxidation. Steam-methane reforming is favored for natural gas as a feedstock, and partial oxidation is reserved primarily for heavier feedstock. The steam-methane reforming process is considered below.

The steam-methane reforming process has been successfully used for hydrogen production for the past 25 yr. In this process, hydrogen is produced by the reaction of steam and methane:



These reactions are carried out over supported nickel catalysts, with conversions being primarily determined by thermodynamic equilibrium. In general, high conversions of methane are favored by high temperatures, low pressures, and high steam gas ratios. Typical temperatures, pressures, and steam gas ratios range from 1000° to 1120°K (1340° to 1556°F), 1.7 to 3.4 MN/m² (247 to 493 psia) and 3:1 to 6:1, respectively. Methane conversions range from 65% to 85%, and about 25% to 75% of the carbon oxides is converted to CO₂.

Further conversion of CO is accomplished at lower temperatures over iron oxide catalysts which are selective for the water gas shift reaction

(Reaction 2). Residual CO_2 can be removed by absorption in hot potassium carbonate solutions, monoethanol amine solutions, or various organic solvents. Subsequent reduction of carbon oxides to trace levels can be accomplished by methanation. Residual methane in the product hydrogen can be reduced by adsorption or cryogenic separation.

The steam methane reforming process requires substantial quantities of energy, with approximately half supplied as high-temperature heat [over 811°K (1000°F)] for the reforming reaction. Energy requirements are typically supplied by the combustion of fuel gas. An alternate approach is to use nuclear heat supplied by an HTGR. The use of nuclear heat allows the substitution of nuclear fuel for a substantial fraction of the gas feedstock required in conventional reforming. Both cases are discussed and compared below.

For the purposes of this technological assessment, the following process steps and process conditions were selected:

1. Production capacity (100% on-stream capacity). $542 \text{ m}^3/\text{sec}$ (1656 MMscfd) at 273°K (32°F); 0.101 MN/m^2 (14.7 psia); 88 mole % H_2 ; product pressure = 10.3 MN/m^2 (1500 psia).
2. Heat source. Case 1: fossil fuel, reformer gas feedstock; case 2: nuclear fuel, HTGR.
3. Reformer process conditions. Process gas outlet temperature = 1006°K (1350°F); process gas outlet pressure = 2.06 MN/m^2 (300 psia); steam carbon mole ratio feed = 4.5 to 1.

Feedstock composition

	<u>Vol %</u>
H ₂	13.3
CH ₄	69.8
C ₂ H ₆	13.7
C ₃ H ₈	1.8
C ₄ H ₁₀	<u>1.4</u>
	100.0

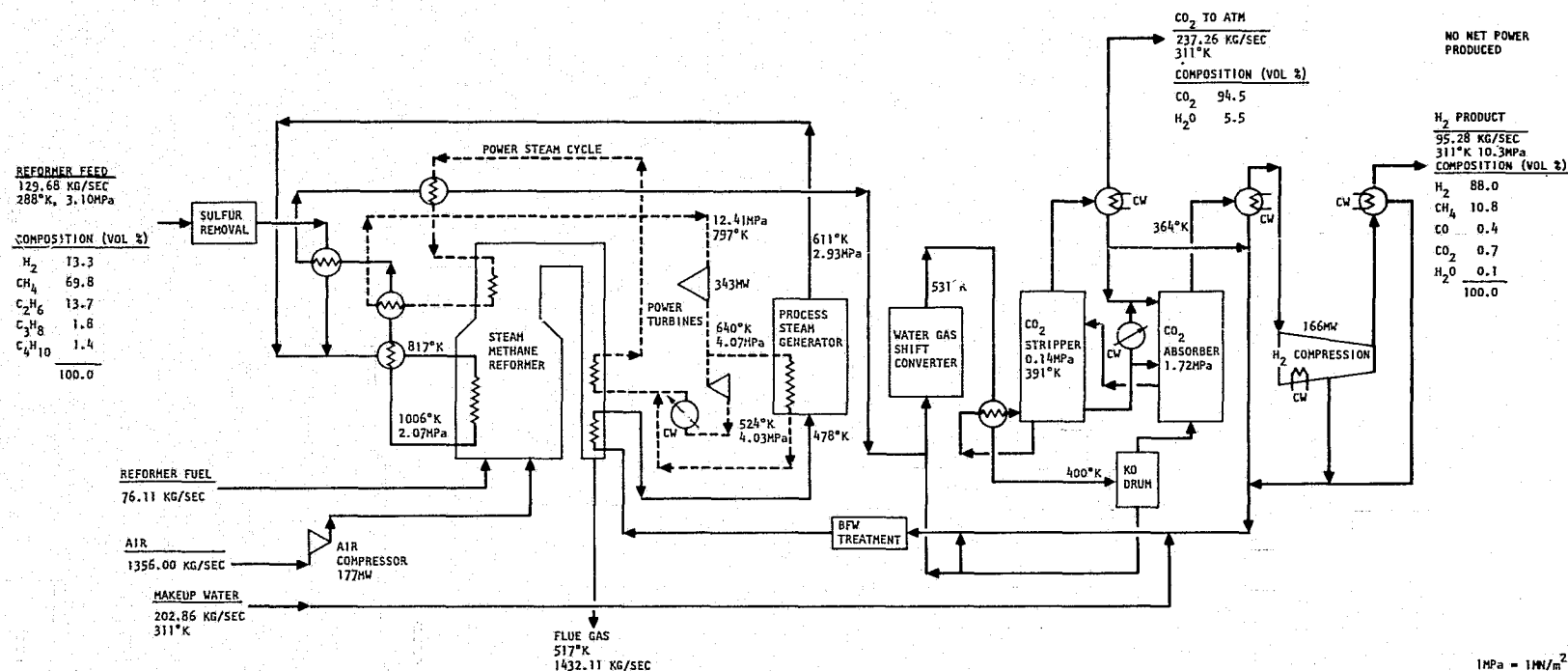
4. Shift conversion. 95% conversion of feed CO.
5. Purification. The volume of CO₂ reduced to 0.6% of the total by absorption in aqueous hot K₂CO₃.

3.1.1. Case I: Fossil Heat Source

3.1.1.1. Process Description. The following process description refers to the flow sheet shown in Fig. 3-1.

Gas feedstock is passed through active carbon beds for trace sulfur removal, mixed with steam, and preheated by heat exchange with the reformer effluent. The feed enters the reformer at 817°K (1010°F) with a steam carbon mole ratio of 4.5 to 1. The steam-methane reforming reaction occurs over a supported nickel catalyst in tubes situated in the radiant zone of the reformer furnace. The heat requirements for the reforming reaction are supplied by the combustion of fuel gas whose composition is identical to that of the reformer feed gas. The reformer effluent leaves at 1006°K (1350°F) and 20.6 MN/m² (300 psia) and has the following composition:

	<u>Vol % (Dry)</u>
H ₂	70.6
CH ₄	9.5
CO	6.9
CO ₂	13.0

Fig. 3-1. H₂ from natural gas steam methane reforming, fossil fired

Hot flue gas leaving the radiant zone of the reformer is used to generate high-pressure steam and preheat boiler feedwater. About 25% of the gross heating value of the fuel gas is discharged to the atmosphere.

Steam generated [12.4 MN/m^2 (1800 psia)] by the hot flue gas is superheated to 797°K (975°F) by heat exchange with the reformer effluent and is used to provide power for the compression of furnace air and hydrogen product. A portion of the steam is condensed in the process steam generator, providing steam for the reformer feed.

The reformer effluent is fed to the shift converters after heat exchange with power steam, reformer feed, and boiler feedwater. Conversion of 95% of the remaining CO is accomplished in two stages, with an intermediate quench with recycled hot condensate. The process gas leaves the shift converters at 531°K (496°F) and is fed to the CO_2 removal system.

A hot potassium carbonate system is used for CO_2 removal. Excess steam in the shift converter effluent is condensed in the CO_2 stripper reboiler and recycled for treatment. The process gas is scrubbed with an aqueous solution of hot potassium carbonate in the carbon dioxide absorber. A split stream absorption system is used, with the upper portion of the absorber operating at a lower temperature to allow a greater reduction in residual CO_2 in the process gas. The process gas leaving the absorber with a residual CO_2 content of 0.6% is cooled, condensing water for recycle, and fed to the hydrogen compressors. The hot potassium carbonate solution is regenerated in the CO_2 stripper. Carbon dioxide and steam leaving the top of the stripper are cooled to recover water for recycle, and the CO_2 -rich gas is vented to the atmosphere. The process gas leaving the CO_2 removal system is compressed in two-stage reciprocating compressors with intercooling. Product hydrogen at 10.3 MN/m^2 (1500 psia) has a purity of 88 mole %.

3.1.1.2. Overall Performance. Plant performance for Case I is summarized in Table 3-1. The total feed gas required is $268 \text{ m}^3/\text{sec}$ (819 MMscfd) at

TABLE 3-1
SUMMARY OF PLANT PERFORMANCE, PROCESS 1, FOSSIL HEAT

	Gross Heating Value		% of Total Input
	(GW)	(10 ⁹ Btu/hr)	
Feed			
Reformer feed gas (169 m ³ /sec)	7.19	24.55	63.0
Reformer fuel (99 m ³ /sec)	4.22	14.41	37.0
Makeup water (203 kg/sec)	--	--	--
Total in	11.41	38.96	100.0
Product			
Hydrogen (542 m ³ /sec, 88 mol % H ₂) ^(a)	8.49	28.99	74.4
Net power produced	--	--	--
Subtotal	8.49	28.99	74.4
Waste heat			
Cooling water	1.81	6.18	15.9
Other	<u>1.11</u>	<u>3.79</u>	<u>9.7</u>
Total out	11.41	38.96	100.0

(a) At 273°K, 0.1013 MN/m².

273°K (32°F) and 0.101 MN/m² (14.7 psia), with fuel accounting for 37% of the total. The gross heating value of the feed is 11.41 MW.

The hydrogen product is produced at a rate of 542 m³/sec (1656 MMscfd) at 273°K (32°F) and 0.101 MN/m² (14.7 psia) 88 mole % H² with a gross heating value of 8.49 MW. Overall efficiency, based on the gross heating value, is 74.4%. Waste heat is primarily attributed to discharge of hot flue gas and cooling water requirements.

3.1.1.3. Duty Cycle. A hydrogen plant of the size considered would use multiple processing trains, affording a plant availability of 100% at fractional capacity. Scheduled outage for a particular processing train would be determined by the catalyst change-out required at a frequency of about once every 2 yr. The time required for catalyst change-out is estimated below on the basis of 20 reformer furnaces per plant, each with 250 reformer tubes. Reformers of this size would be considered large by present day standards.

Catalyst change-out requires the following sequence of events: (1) cooling and purging the reformer; (2) disconnecting the flanges; (3) vacuuming out the catalyst; (4) charging the fresh catalyst by floating it through water; (5) drying the catalyst; (6) reconnecting the flange; and (7) purging and start-up. Cooling and purging the reformer tubes takes one day, based on metallurgical considerations. Catalyst replacement (steps 2 through 4) can be accomplished at a rate of about 1/2 hr/tube, or 5 days for 250 tubes. Catalyst drying and start-up would require an additional 2 days, resulting in a total of 8 days required for catalyst change-out.

Additional scheduled outage is required for retubing reformers at an expected frequency of about once every 5 to 10 yr. The time for retubing the reformer is estimated to be 20 days.

Based on scheduled outage, the availability of a particular processing train would be better than 95%. Unplanned maintenance should not reduce availability below 90%.

3.1.1.4. Status of Process Technology. The steam-methane reforming process has been reliably operated in petroleum refineries and ammonia plants for the past 25 yr. The process is well within the present day technology. Plant capacities, however, range up to 27 MMscfd of hydrogen, for below the 574 MMscfd considered in this report. The large capacity was selected for comparison with hydrogen production from a nuclear heat source with a 3000-MW(t) capacity.

3.1.1.5. Status of Materials Technology. Metallurgical considerations are of particular importance in reformer design and are a major factor in limiting reformer performance. The primary component in terms of quantity is the reformer tube. A reformer tube history consists of steady-state creep under internal pressure at temperatures of 1000° to 1250°K (1340° to 1790°F). In addition, cyclic variations and nonuniform heating occur during start-up and shutdown sequences.

Cast iron chromium-nickel heat-resistant alloys, such as HK-40, are commonly used. Designs are based on a limited reformer life, which can typically range from 5 to 10 yr. A history of satisfactory operating experience has been demonstrated. Further improvements in tubing materials could lead to significant improvements in the process.

3.1.1.6. Resource and Environmental Considerations. The principal resource consideration for the steam-methane reforming process is the availability of natural gas. As natural gas becomes scarce, associated higher prices could render the process uneconomical compared with other processes. Higher molecular weight hydrocarbons, up to naphthas, are also suitable feedstock for the process and are extensively used in Europe, where shortages of natural gas currently exist. However, there will also be shortages of these resources in the foreseeable future. To date, the process has not been demonstrated using a hydrocarbon feedstock heavier than naphtha. Synthetic natural gas obtained from coal, oil shale, tar sand, or petroleum may be a possible future feedstock for the process. This approach is considered in detail in Section 3.2.

Water requirements for the steam-methane process include makeup water at 203 kg/sec (3220 gpm) for the reformer feed and cooling water. The cooling water heat load is 1810 MW, which, if supplied by wet cooling towers, will require 600 kg/sec (9520 gpm) of water.

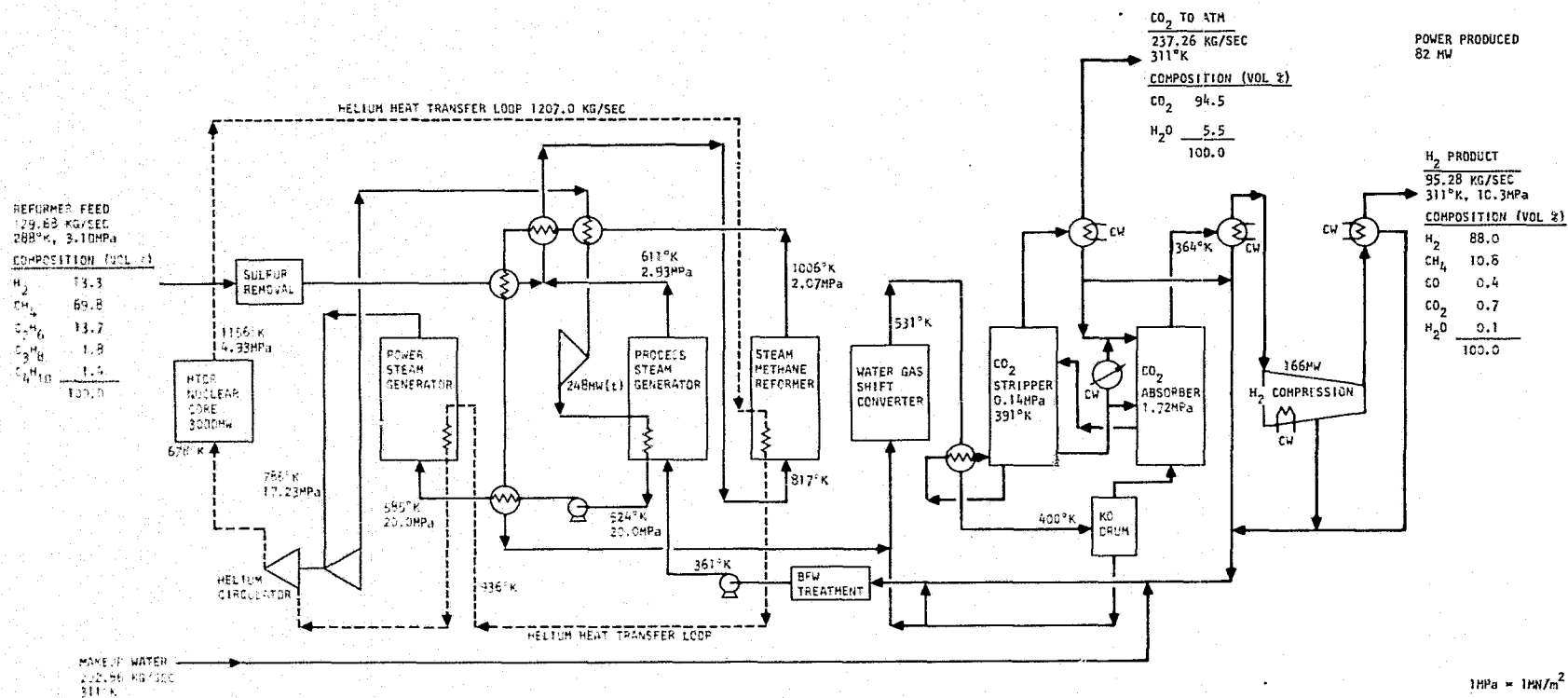
The process has minimal waste disposal problems. Gas feedstock is assumed to be sufficiently free of pollutants to allow direct discharge of flue gas to the atmosphere.

3.1.2. Case II: Nuclear Heat

Steam-methane reforming using nuclear heat replaces the fossil-fired reformer furnace with an HTGR. A convectively heated reformer is required, with heat being supplied by a stream of hot helium. Process steam is produced in a similar manner. The remaining process steps are similar to those for Case I.

3.1.2.1. Process Description. The following process description refers to the flow sheet shown in Fig. 3-2.

Process power requirements are entirely supplied by an HTGR. Helium, which serves as the primary reactor coolant, carries heat from the nuclear core to the reformer and power steam generator. A PCRV shields and contains the nuclear core and primary coolant system, including reformers, steam generators, and helium circulators. Helium entering the nuclear core at 678°K (760°F) and 5.0 MN/m² (725 psia) is heated to 1156°K (1620°F) by a high-enrichment uranium-thorium fuel. The hot helium supplies heat to the reformers, leaving at 941°K (1235°F). The helium steam then transfers additional heat to the power steam generators. The helium is compressed by the helium circulators and returned to the core. The thermal power of the HTGR core is distributed as follows:

Fig. 3-2. H₂ from natural gas steam methane reforming, nuclear heated

<u>Component</u>	<u>MW</u>	<u>MMBtu/hr</u>
Reformer	1,380	4,711
Steam generator	<u>1,700</u>	<u>5,804</u>
Subtotal	3,080	10,515
Helium circulation	<u>-80</u>	<u>-273</u>
Total HTGR thermal power	3,000	10,242

The flow sheet for this process is shown in Fig. 3-2; reformer feed gas is passed through zinc oxide beds for sulfur removal, mixed with steam, and preheated to 817°K (1010°F). The process gas leaves the reformer at 1006°K (1350°F) and 2.06 MN/m² (300 psia). The reformer effluent is cooled by heat exchange with steam, reformer feed, and boiler feedwater. The remaining process steps are identical to those described in Section 3.1 for hydrogen from gas using fossil heat (Case I).

Power steam is generated by the HTGR at 17.2 MN/m² (2500 psia) and 786°K (955°F). The high-pressure steam is used to provide power for helium circulation and is then reheated to 797°K (975°F) by the reformer effluent to provide process power primarily for hydrogen compression. The steam is condensed in the process steam generator, providing steam required for reformer feed. The boiler feedwater is recycled to the power steam generator after it is preheated by the reformer effluent.

3.1.2.2. Overall Performance. Plant performance for Case II is summarized in Table 3-2. The total feed gas required is 169 m³/sec at 273°K (32°F) and 0.101 MN/m² (14.7 psia) with a gross heating value of 7.19 GW. Process heat and power requirements are supplied by a 3000-MW(t) HTGR.

The hydrogen product is produced at a rate of 542 m³/sec (1656 MMscfd) at 273°K (32°F) and 0.101 MN/m² (14.7 psia) 88 mole % H₂ with a gross heating value of 8.49 GW. Overall efficiency is 84%. Waste heat is principally attributed to cooling water requirements.

TABLE 3-2
SUMMARY OF PLANT PERFORMANCE, PROCESS 1, NUCLEAR HEAT

	Gross Heating Value		% of Total Input
	(GW)	(10 ⁹ Btu/hr)	
Feed			
Reformer feed gas (169 m ³ /sec)	7.19	24.55	70.6
HTGR	3.00	10.24	30.4
Makeup water (203 kg/sec)	--	--	--
Total in	10.19	34.79	100.0
Product			
Hydrogen (542 m ³ /sec, 88 mol % H ₂) ^(a)	8.49	28.99	83.3
Power (electric)	0.07	0.24	0.7
Subtotal	8.56	29.23	84.0
Waste heat			
Cooling water	1.57	5.36	15.4
Other	0.06	0.20	0.6
Total out	10.19	34.79	100.0

(a) At 273°K, 0.1013 MN/m².

3.1.2.3. Duty Cycle. Plant availability studies have previously been reported (Ref. 3-1). Refueling of the nuclear core determines the time required for scheduled outages. The standard 3000-MW(t) HTGR is designed for a 4-yr fuel cycle with refueling of one-quarter of the core every year; refueling requires 16 days. Additional unscheduled maintenance is not expected to reduce availability below 90%.

The plant capacity factor, defined as the ratio of actual hydrogen production to rated production capacity, is limited by the nuclear fuel life. For the standard 3000-MW(t) HTGR, full-power years on a 4-yr cycle are 3.2, affording a plant capacity factor of 0.80.

Preliminary designs for nuclear heated reformers are for a design life of 30 yr, or the expected life of the plant. On this basis, periodic retubing of the reformer is not required. For a design life of less than 30 yr, reformers would be replaced with spares as required. This operation requires approximately 95 hr. It would be necessary for the manufacturer to provide a central service facility to accomplish retubing.

3.1.2.4. Status of Process Technology. The HTGR for electric power production has been under development at GA since 1957. The development effort included the 40-MW(e) prototype plant at Peach Bottom (Pennsylvania), which was successfully operated from 1968 to 1974. The 330-MW(e) HTGR at Fort St. Vrain (Colorado) is scheduled for commercial operation in 1975. Over the past 12 yr, a substantial research and development effort has been directed toward the large 3000-MW(t) HTGR, and two electric utilities have made the decision to build large HTGR plants.

The use of the HTGR for hydrogen production by steam-methane reforming requires design modifications of several components and a limited amount of new development effort. To a great extent, however, the basic technology of the present day HTGR remains applicable.

The production of hydrogen by reforming natural gas at 1006°K (1350°F) will require a helium temperature of 1156°K compared with a temperature of

1033°K (1400°F) in current electric power plants. Two approaches are available for increased helium temperature: (1) modification of the current fuel block design, which would enable the required increase in helium temperature without the associated increase in fuel temperature, and (2) further developments in fuel technology, which would allow higher helium temperatures to be obtained through increases in fuel temperatures.

The use of convectively heated reformers is a significant departure from conventional reforming practices. However, preliminary studies have indicated that convective heating of reformers with hot helium is feasible with reaction conditions similar to those used in conventional reforming. Results of full-scale, single-tube reformer tests conducted by Kernforschungsanlage in Germany have further demonstrated the feasibility of nuclear-heated reformers.

3.1.2.5. Status of Materials Technology. As with conventional reforming, tube temperatures in reformers approach the limits of those for conventional metallic materials. However, for the process conditions considered in this report, tube temperatures are lower than those typically used in fired reformers. Preliminary designs based on a 30-yr reformer life have indicated that the cast iron-chromium-nickel heat-resistant alloys (e.g., HK-40) and wrought iron-chromium-nickel alloys (e.g., Incoloy 800) are candidate materials. Materials technology applicable to nuclear heated reformers is discussed in detail in Ref. 3-1.

Increased helium temperatures in the PCRV require reconsideration of thermal barrier materials. Present thermal barrier designs for the HTGR electric power plants consist of fibrous insulation contained by metal cover plates. For a design with a core outlet temperature of 1156°K (1620°F), the temperature limits of conventional metallic materials will be exceeded under accident conditions. Molybdenum cover plates or an all-ceramic thermal barrier may be required in hot helium environments. Materials currently used within the nuclear core are capable of service at the temperatures considered.

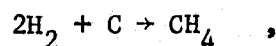
substitution of nuclear fuel for a portion of the natural gas feedstock required in the conventional fossil-heated reforming process. As shortages of natural gas develop, nuclear heated reforming should become increasingly attractive from the standpoints of conservation and economics. Severe shortages of natural gas would require the use of synthetic natural gas as a feedstock. This process is considered in Section 8.

Water requirements include makeup water for the reformer feed at 203 kg/sec and cooling water. The cooling water heat load is 1570 MW, which, if supplied by wet cooling towers, would require 680 kg/sec of water.

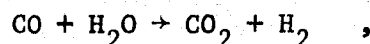
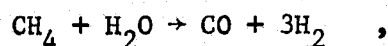
3.2. HYDROGEN FROM COAL (PROCESS 2)

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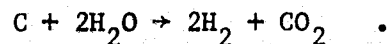
Hydrogen can be produced from coal by hydrogasification,



followed by steam hydrocarbon reforming,



giving the overall reaction



The process described below uses coal solubilizing techniques to produce coal liquids which are hydrogasified to produce synthetic pipeline gas. Hydrogen is produced by steam reforming of the synthetic gas using a nuclear heat source. The process is essentially the same as the nuclear coal solution gasification process reported in Ref. 3-2, modified to give hydrogen as the principal product.

3.2.1. Process Description

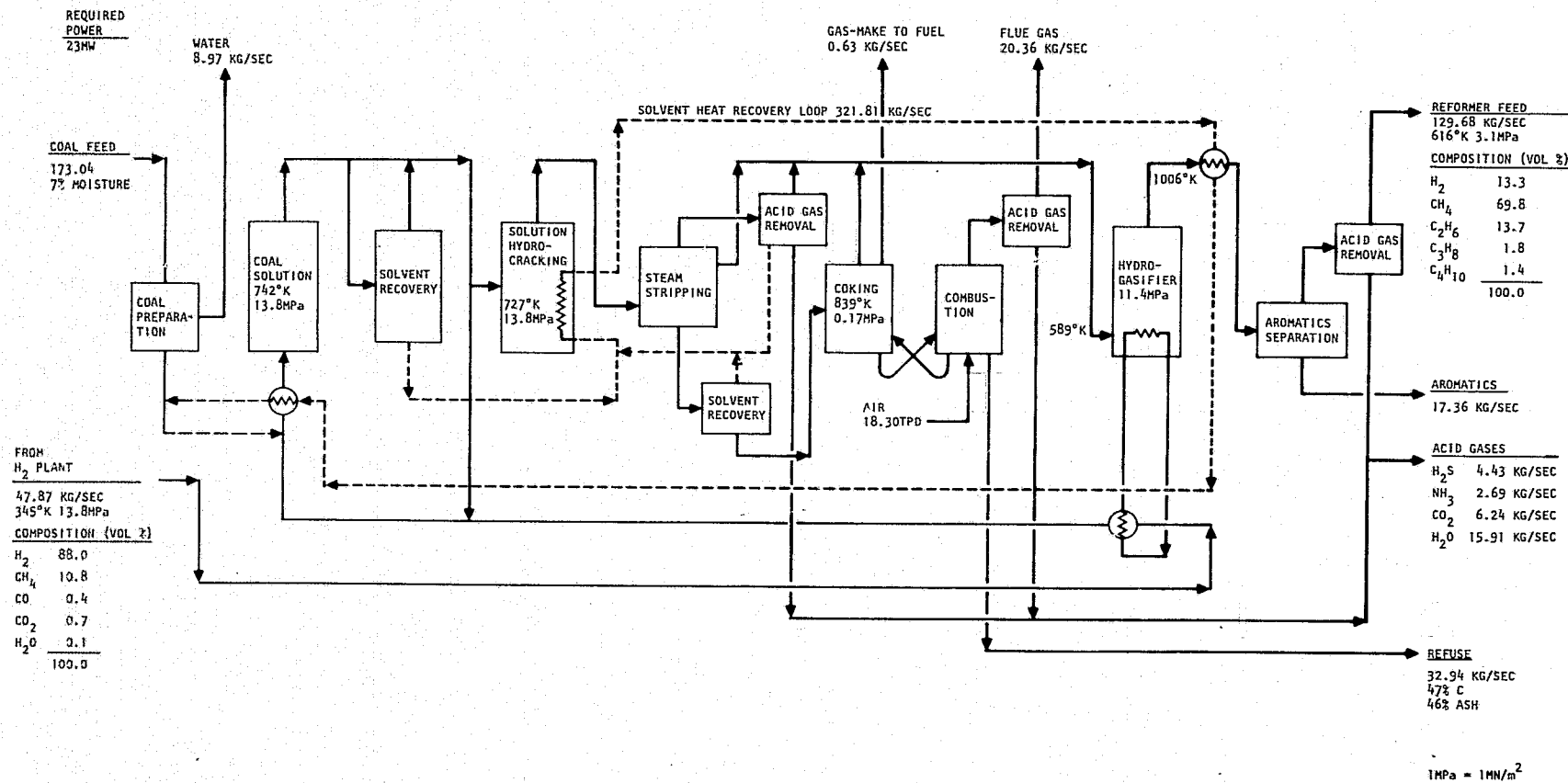
A flow sheet for the hydrogasification process is shown in Figs. 3-3 and 3-4. The hydrogen plant shown in Fig. 3-4 is essentially the same as that described in Section 3.1. The coal gasification plant attached to the front end of the hydrogen plant is shown in block diagram form in Fig. 3-4 with only major heat exchange indicated.

3.2.1.1. Coal Preparation and Solubilization. Coal is prepared for solubilization by pulverizing and drying with hot CO_2 . The pulverized coal is slurried with 2 kg of recycle solvent per kg of coal. The slurry is preheated by heat exchange with recycle solvent and fed to the coal solution reactor along with half of the hydrogen required for the process.

In the coal solution reactor, coal molecules are depolymerized under high-pressure hydrogen at 13.7 MN/m^2 (2000 psi) and 742°K (875°F). Nearly all the coal is converted to distillable coal liquids consisting primarily of complex aromatics. Approximately 10% of the hydrogen is consumed.

A portion of the coal solution reactor product is sent to the solvent recovery section, and the remainder is directly fed to the solution hydrocracking reactor. Approximately 25% of the solvent is recovered and recycled through the solvent heat recovery loop.

3.2.1.2. Solution Hydrocracking. The molecular weight of the coal liquids is further reduced by catalytic solution hydrocracking. Hydrocracking conditions of 728°K (850°F) and 13.7 MN/m^2 (2000 psi) were selected on the basis that these conditions have previously been demonstrated. About 78% of the hydrocarbon feed (excluding solvent) is converted to hydrocarbon gases and distillate suitable for hydrogasifier feed. Heat released by the hydrocracking reactions is removed by heat exchange with recycle solvent. Hydrocarbon gases and distillate suitable for hydrogasifier feed are recovered by flash separation and atmospheric steam stripping. A portion

Fig. 3-3. H₂ from coal liquefaction

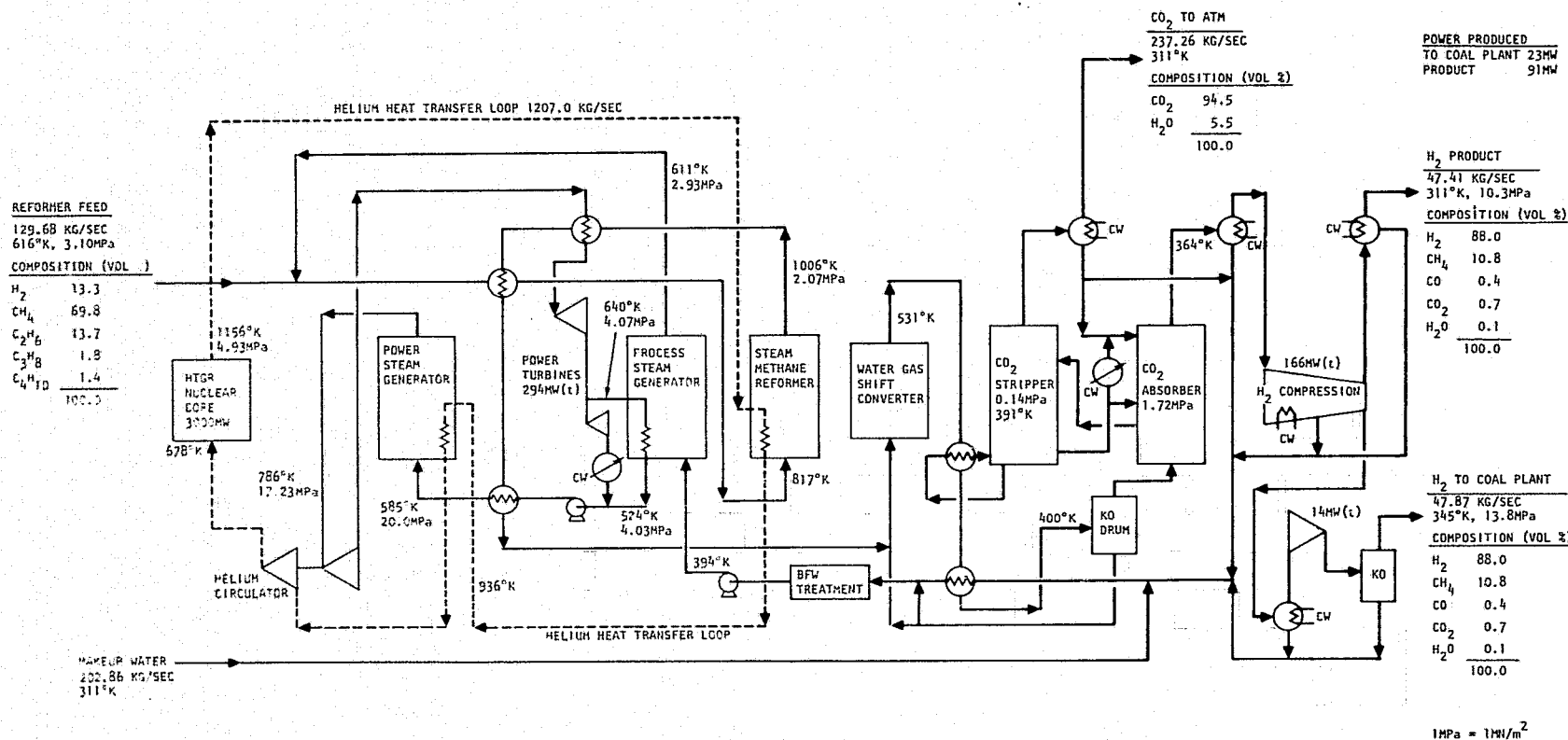


Fig. 3-4. Hydrogen from coal by hydrogasification steam methane reforming and hydrogen purification

of the acid gases is removed at this point. Stripper bottoms consist of solvent and heavy coal liquids. Solvent is recovered by vacuum flash separation, and the heavy coal liquids are sent to the fluid bed coker.

3.2.1.3. Fluid Bed Coking. Fluid bed coking is carried out at 811°K (1000°F) and essentially atmospheric pressure. The reactor bed is fluidized by product vapors and steam. Approximately half of the heavy coal liquids is converted to hydrogasifier feed, and the remainder is discharged as refuse consisting of 47% carbon and 46% ash. A small portion of coke is burned in the fluid bed combustion zone, providing heat and power for the coking process. Three thousand tons of refuse with a gross heating value of 590 MW are discharged per day, representing about 11% of the gross heating value of the coal.

Various methods are available for obtaining greater utilization of the carbon in the feed coal. Increased conversion of the coal to hydrocarbon liquids suitable for hydrogasifier feed is one possibility. However, the extent to which this approach can be pursued cannot be determined without further experimentation. Another approach is direct gasification of the char with steam and oxygen or hydrogen. Partial oxidation with mixtures of steam and oxygen would allow conversion of 60% to 80% of the remaining carbon in the refuse to hydrogen and carbon oxides, which could be mixed with the reformer product gas for purification. Hydrogasification would yield a product gas rich in methane, providing additional reformer feed gas.

The carbon in the refuse could be used to generate process power by either combustion or partial oxidation with steam and air to produce a low-Btu gas for fuel. This additional source of process power would allow a greater fraction of HTGR power to be used for reforming, which could be accomplished through increases in core inlet temperature and helium flow rate.

An economic analysis would be required to justify modification of the process to reduce the carbon content of the refuse. Such a study is

considered beyond the scope of this technology assessment. However, methods are available for a threefold to fourfold reduction of the carbon content of the refuse, affording increases in carbon utilization from 89% to 97%.

3.2.1.4. Hydrogasification. Hydrocarbon gases, including hydrogen, and hydrocarbon liquids are fed to the hydrogasifier at 589°K (600°F). Hydrogasification consists primarily of the following reactions: (1) conversion of complex aromatics to simple aromatics and C₁ to C₄ hydrocarbons and (2) hydrogenation and decomposition of simple aromatics, yielding C₁ to C₄ hydrocarbons. The hydrocarbons are heated to 1006°K (1350°F) by heat released by the hydrogasification reactions. Additional heat is removed by vaporization of boiler feedwater at high pressure, which is in turn used to preheat hydrogen feed to the coal solution reactor and the solution hydrocracking reactor. The product gases are cooled by heat exchange with recycle solvent before final purification.

Conversions in the hydrogasifier are limited by reaction kinetics. Conversion of complex aromatics to simple aromatics and light hydrocarbons occurs more rapidly than subsequent conversion of simple aromatics to hydrocarbon gases. The hydrogasifier is sized for incomplete conversion of light aromatics, and the unconverted fraction is recovered as byproduct.

The aromatics are separated from the hydrogasifier effluent by lean oil absorption. The lean oil is regenerated in a stripper, and aromatics are recovered. Acid gases and residual aromatics are removed by absorption in methanol. Ammonia, hydrogen sulfide, and carbon dioxide are also separated. Hydrogen sulfide is sent to the sulfur recovery plant, and CO₂ is vented to the atmosphere. Aqueous ammonia is produced as byproduct. The product gas rich in methane and ethane is reheated to 617°K (650°F) after trace sulfur removal and is used as reformer feedstock for the hydrogen plant.

3.2.1.5. Hydrogen Production and Power Generation. Hydrogen production and power generation are essentially the same as for hydrogen from gas using a nuclear heat source, as described in Section 3.1.2.

3.2.2. Overall Performance

A summary of overall plant performance is given in Table 3-3. Coal feed is required at a rate of 173 kg/sec with a gross heating value of 5.0 GW. Process heat and power requirements are supplied by a 3000-MW HTGR.

Product containing 88 mole % hydrogen is produced at a rate of 269 m³/sec (822 MMscfd) at 273°K (32°F) and 0.101 MN/m² (14.7 psia) with a gross heating value of 4.08 GW. Aromatics and char (47% carbon) are produced as byproducts. The aromatics cannot be sold in the present chemical market because their quality is not sufficiently high. However, they are very low in sulfur and should be valuable as fuel. Upgrading the aromatics for sale as petrochemical feedstock is possible. Overall efficiency, including char and aromatics, is 70.2%.

3.2.3. Duty Cycle

Plant availability is determined by the refueling time required for the HTGR. As discussed in Section 3.1.2.3, this limits availability to 95%. The plant capacity factor, defined as the ratio of actual hydrogen produced to rated capacity, is limited by the life of the nuclear fuel. The plant capacity factor is 0.80.

3.2.4. Status of Process Technology

3.2.4.1. Coal Solubilization. Gulf Research and Development Corporation is currently developing a catalytic coal liquids process for the production of sulfur-free boiler fuels. Conversion of over 90% of the coal to liquids has been demonstrated in a 3 lb/day pilot plant. Start-up of a 3 ton/day pilot plant is under way.

TABLE 3-3
SUMMARY OF OVERALL PERFORMANCE, PROCESS 2, HYDROGEN FROM COAL

	Gross Heating Value		% of Total Input
	(GW)	(10 ⁹ Btu/hr)	
Input			
Coal (173 kg/sec)	4.99	17.04	62.5
HTGR	3.00	10.24	37.5
Water (203 kg/sec)	--	--	--
Total in	7.99	27.28	100.0
Product			
Hydrogen (269 m ³ /sec, 88 mol % H ₂) ^(a)	4.19	14.31	52.45
Aromatics	0.73	2.50	9.2
Char	0.59	2.02	7.4
Power (electric)	0.09	0.31	1.1
Subtotal	5.60	19.14	70.15
Waste heat			
Cooling water	2.07	7.05	25.85
Other (including acid gases)	0.32	1.09	4.0
Total out	7.99	27.28	100.0

^(a)At 273°K, 0.1013 MN/m².

3.2.4.2. Solution Hydrocracking. Hydrocracking processes have been directed toward the production of liquid products, limiting temperatures to between 800° and 850°F. Consequently, highly active catalysts and high hydrogen pressures are required for favorable kinetics. The catalytic process is reasonably well developed in petroleum refining. However, hydrocracking of coal liquids has not been demonstrated to date.

3.2.4.3. Fluid Bed Coking. Fluid bed coking is currently carried out in petroleum refineries using vacuum tower bottoms as feedstock. This material is probably very similar chemically to the coker feed used in this process. However, normal feedstocks contain only small quantities of ash, and the principal area of uncertainty pertains to ash distribution. Alternate solutions to the problem of ash separation exist, including filtration and partial oxidation.

3.2.4.4. Hydrogasification. There is a significant amount of literature on catalytic hydrocracking of aromatic and paraffinic liquids. However, only a limited amount of data is available on strictly thermal reactions in the presence of hydrogen. One area of uncertainty is the extent of coke formation from thermal cracking and the associated operating conditions required to eliminate coke formation. However, it appears that distillate feedstocks can be gasified with very low productions of net carbon. Initial experimental results at low pressures have been encouraging.

3.2.4.5. Hydrogen Plant. The status of the process technology for production of hydrogen using a nuclear heat source is discussed in Section 3.1.

3.2.5. Status of Materials Technology

The status of materials technology for the hydrogen plant is discussed in Section 3.1. For the remaining portions of the plant, temperature and pressure conditions are such that conventional materials are adequate.

3.2.6. Environmental and Resource Considerations

The solubilization of coal liquids has been demonstrated on a pilot plant scale for a wide variety of coals. Coal reserves in the United States are sufficient to ensure the availability of coal feed in the foreseeable future. Environmental considerations associated with mining the coal vary considerably according to the type of mining used and are considered beyond the scope of this report.

About 80% of the sulfur in the coal is recovered as H_2S , which requires conversion to solid sulfur by a Claus oven system before ultimate disposal. Sulfur dioxide present in the flue gas leaving the fluid bed combustor in the coker is also sent to the Claus oven system. Because less than 1% of the coal is burned, no problems are anticipated in venting nitrogen oxides to the atmosphere.

Water generated from oxygen in the coal during coal processing contains small percentages of phenolics, but bacteriological treatment of wastewater reduces the phenolics to acceptably low levels. Water requirements include makeup water for reformer feed at 203 kg/sec (3220 gpm) and cooling water. The cooling water heat duty is 2180 MW, which, if supplied by wet cooling towers, requires 725 kg/sec (10,300 gpm) of water.

3.3. PRODUCTION OF HYDROGEN BY DIRECT GASIFICATION OF COAL WITH STEAM USING NUCLEAR HEAT

N76-15577

3.3.1. Introduction

Steam gasification of coal is based on the reaction of carbon with steam according to the reaction equation



which shows that this process is highly endothermic. In all conventional plants, the fossil fuel of the gasifier is also the source of reaction heat and all other energy requirements. Obviously, available coal reserves can be saved by using process heat from an HTGR for steam gasification and the production of necessary process steam. In this case, the coal is only used as raw material for the produced gas. Additional advantages are production of smaller amounts of CO_2 in the gasifier and, depending on the coal price, eventual lower gas production costs.

In order to realize reasonable gasification rates and hence throughputs of the gasifier, the helium outlet temperature of the HTGR, which determines the maximum temperature in the gasifier, has to be as high as possible, with a minimum value of about 1220°K (1745°F). This lower limit depends on the type of coal to be gasified and the heat transfer characteristics of the gasifier. In general, it can be stated that the higher the helium outlet temperature of the HTGR, the higher the overall efficiency of the plant, the reason being that the distribution of the thermal output of the reactor between gas and power generation shifts to the gas side at higher helium temperatures. Because each process, other than process heat, also needs power for steam generation, the optional distribution of the two uses must be found. In this technological assessment, the output of the HTGR is used for the gasification process, shift conversion, and purification of the product gas, and 406 MW of electric power are generated. This study is based on the use of a 3000-MW(t) process heat HTGR giving an outlet helium temperature of 1366°K (2000°F).

No basic data are available for the gasification of Oklahoma coal in a fluidized bed gasifier at expected temperatures and pressures, and the kinetics of the heterogeneous gasification reactions are unknown. The results of the laboratory and bench-scale experiments carried out in Germany with Hagen coal have been used as a guide for this assessment (Refs. 3-3, 3-4).

The most important conclusion that can be drawn from GA development work for HTGR process heat applications and from published results of the development work under way in Germany is that a process such as the one described in this section is feasible and the problems related to the high helium outlet temperature of the reactor and the design of the gas generator can be solved.

The GA conception of the plant design is described in Section 3.3.2. In this plant, 117 kg/sec of Oklahoma coal is gasified to produce 247 m³/sec (796 MMgcf/d) of product gas containing 82.7% H₂ at 10.3×10^6 N/m² (1493 psia).

3.3.2. Process Description

An intermediate helium loop has been selected to transfer heat from the reactor to the gasifier. The operation of the fluidized bed of the gasifier is not felt to be suitable for operation within the reactor because of potential operating difficulty and size. Figure 3-5 shows how the energy is transferred and divided over the various heat exchangers. The heat is discharged from the reactor by a helium circuit [the temperature of the helium leaving the reactor is 1366°K (2000°F)], transferred to the secondary helium loop in the intermediate heat exchanger, where 2564 Gcal/hr are lost, after which the helium, at a temperature of 841°K (~1060°F), returns to the reactor. The helium flow rate in the primary loop is equal to 1093 kg/sec, and the pressure is equal to 5×10^6 N/m² (725 psia).

In order to keep the heat exchanger area within reasonable dimensions, a minimum temperature difference of 50°K (90°F) between the gasification temperature and the helium outlet temperature of the reactor has been assumed. Hence, the helium outlet temperature of the intermediate heat exchanger of the secondary loop is equal to 1316°K (~1910°F). Based on the results of the experiments described in Ref. 3-3, the outlet

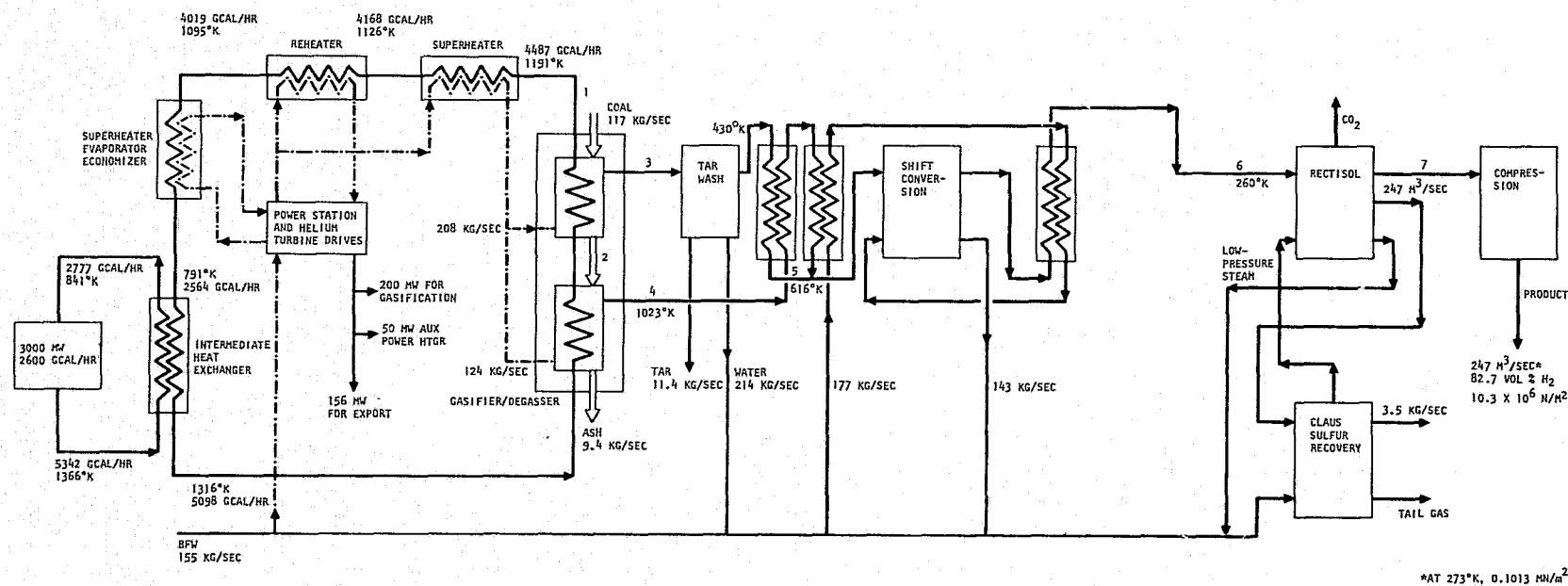


Fig. 3-5. Hydrogen production using steam coal gasification

temperature of the degasser helium is limited to 1093°K (1508°F) in order to achieve reasonable gasification rates. In the present process the outlet temperature of the degasser is equal to 1191°K (1684°F), so the energy input in the degasser and the fluidized bed gasifier is equal to 611 Gcal/hr, or 23.8% of the thermal heat produced by the reactor.

Next, the helium flows through the superheater, where 18.4 kmole/sec steam are heated to 1073°K (1472°F) and 319 Gcal/hr are lost, after which 149 Gcal/hr are transferred in the reheater and 1455 Gcal/hr in the superheater, economizer, and evaporator. The helium inlet temperature of the intermediate heat exchanger is equal to 791°K (964°F).

According to Ref. 3-3, a heat exchanging area of 80 m² (860 ft²) is needed to gasify 1 ton of coal in 1 hr. Hence, the required heat exchanging area is approximately 35,000 m² (376,600 ft²) or nine 4000-m² (43,040-ft²) units. Each gasifier is a cylinder in a horizontal position and is about 30 m (98.4 ft) long and 7 m (23.0 ft) in diameter. The fluidized bed is heated by a helium-to-fluidized-bed heat exchanger. The effective volume of the fluidized bed in the gas generator is approximately 320 m³ (11,300 ft³), and the fluidizing agent is superheated steam at a temperature of 1073°K (1472°F).

Oklahoma coal (see Table 3-4) is fed to the degasser at a rate of 117 kg/sec, and the degassed coal and ash go to the gasifier (stream 2). After tar wash, the effluent gas stream of the degasser (stream 3) is combined with the product gas from the gasifier (stream 4) and steam at 5 x 10⁶ N/m² (725 psi) and 620°K (660°F) and fed to the shift converter (see Table 3-5) where the carbon monoxide is transferred into CO₂ according to the reaction equation



TABLE 3-4
COAL GASIFICATION WITH STEAM USING
NUCLEAR HEAT MATERIAL BALANCE

	Stream No. 1 (117 kg/sec)			Stream No. 2 (63.5 kg/sec)		
	wt %	kmole/sec	kg/sec	wt %	kmole/sec	kg/sec
H ₂ O	6.4	0.414	7.47	--	--	--
Ash	8.0	--	9.33	14.7	--	9.33
C _{vol}	21.1	2.053	24.60	--	--	--
H ₂	5.0	2.917	5.83	--	--	--
O ₂	7.7	0.280	8.97	--	--	--
N ₂	1.6	0.064	1.86	--	--	--
C _{fixed}	46.4	4.510	54.14	85.3	4.51	54.14
S	3.8	0.139	4.55	--	--	--

TABLE 3-5
COAL GASIFICATION WITH STEAM USING NUCLEAR HEAT MATERIAL BALANCE

	Stream No.									
	$(78.53 \text{ m}^3/\text{sec})^{(a)}$		$(235.6 \text{ m}^3/\text{sec})^{(a)}$		$(281.8 \text{ m}^3/\text{sec})^{(a)}$		$(358.4 \text{ m}^3/\text{sec})^{(a)}$		$(247.1 \text{ m}^3/\text{sec})^{(a)}$	
	Mol %	kmole/sec	Mol %	kmole/sec	Mol %	kmole/sec	Mol %	kmole/sec	Mol %	kmole/sec
H ₂	26.03	0.913	57.03	6.00	54.92	6.91	57.01	9.12	82.69	9.12
CO	5.65	0.198	20.44	2.15	18.60	2.34	0.87	0.14	1.27	0.14
CO ₂	0.66	0.023	20.44	2.15	17.25	2.17	27.31	4.37	0.91	0.1
CH ₄	17.46	0.612	2.09	0.22	6.60	0.83	5.18	0.83	7.5	0.83
C _n H _m	5.65	0.198	--	--	1.59	0.20	1.25	0.20	1.81	0.20
N ₂	1.83	0.064	--	--	0.56	0.07	4.00	0.64	5.80	0.64
H ₂ S	2.00	0.070	--	--	0.48	0.06	4.4	0.70	--	--
H ₂ O	11.81	0.414	--	--	--	--	--	--	--	--
Tar	28.91	1.014	--	--	--	--	--	--	--	--
Total	100.00	3.506	100.00	10.52	100.00	12.58	100.00	16.00	100.00	11.03

(a) At 273°K, 0.1013 MN/m².

The steam/CO molar ratio is 4.2/1, and the reaction takes place over a special high-temperature shift catalyst. Approximately 95% of the CO is converted (see Table 3-5, stream 6). High-temperature catalysts have been chosen because they tolerate H_2S in the gasified stream, thus avoiding H_2S removal before shift conversion. About seven to eight shift converters are required, each 3.5 m (11.5 ft) in diameter, 18 m (59 ft) high, and containing 15 m (49.2 ft) [approximately 145 m^3 (5120 ft^3)] of catalyst. After waste heat recovery, the product gas is fed to the Rectisol gas purification units, where all remaining sulfur compounds and practically all CO_2 are removed by a countercurrent wash with methanol at a temperature of $\sim 200^\circ K$ ($\sim 100^\circ F$) (see Table 3-5, stream 7). For $355\text{ m}^3/\text{sec}$ ($13,252\text{ ft}^3/\text{sec}$) of product gas, 22 Rectisol purification units and 7 methanol regeneration sections are needed. The acid gas stream of the Rectisol units is fed to the Claus unit for sulfur recovery. Oklahoma coal can be gasified with the equipment described at a rate of 117 kg/sec, producing $247\text{ m}^3/\text{sec}$ ($9,220\text{ ft}^3/\text{sec}$) of product gas containing 82.7 mole % H_2 ($\sim 18.3\text{ kg/sec}$ of 100%-pure H_2). Hence, to produce 1 kg/sec of 100%-pure H_2 using the present process, 6.4 kg/sec of Oklahoma coal are needed. If the Lurgi coal gasification route described in Section 3.4 is used, 14.5 kg/sec of Oklahoma coal, i.e., 226% more, is needed for the same amount of H_2 . Overall efficiency is defined as

$$\eta = \frac{\text{Higher heating value of product gas (kcal/sec)} + \text{generated export power (kcal/sec)}}{\text{Higher heating value of coal input (kcal/sec)} + \text{nuclear heat input (kcal/sec)}} \times 100,$$

which is equal to 59.7%. This overall efficiency figure is a very conservative value. Higher values can probably be realized by improving the heat exchangers and work heat recovery, thus reducing the percentage of nuclear heat used for process steam generation, and by reducing the generated electric power to the exact quantity needed by the process.

3.3.3. Overall Performance

The overall plant performance is summarized as shown in Table 3-6. The overall thermal efficiency of the plant is equal to 59.7%. Twenty-four percent of the available heat is used in the degasser-gasifier combination, 12.4% is used for the superheater, 5.8% is used for the reheater, and 56.7% is used for the superheater, economizer, and vaporizer. Recovered heat from the waste heat recovery section and the Claus sulfur recovery unit is used to produce steam for shift conversion and stripping in the Rectisol units. A breakdown of the plant energy distribution is shown in Table 3-6.

3.3.4. Duty Cycle

Plant availability for the nuclear process heat plant is defined as the percent of total time in a given period during which the plant produces hydrogen. The plant capacity factor is defined as the percent of rated hydrogen production capacity which the plant actually produced. Table 3-7 outlines the total number of days of scheduled outage due to refueling. Catalyst replacement can be carried out during refueling periods. It is assumed that the material problems encountered in high-temperature operation can be solved in such a way so that no down time for unscheduled maintenance due to temperature-related failures will be necessary.

3.3.5. Status of Process Technology

In order to realize this project on an industrial scale, it is necessary to carry out the development of

1. High-temperature nuclear reactors with a helium outlet temperature of 1370°K (2000°F).
2. Special heat transfer equipment.
3. Gas generators.

TABLE 3-6
SUMMARY OF OVERALL PERFORMANCE, PROCESS 3, DIRECT GASIFICATION
OF COAL WITH STEAM USING NUCLEAR HEAT

	Gross Heating Value		% of Total Input
	(GW)	(10 ⁹ Btu/hr)	
Input			
Coal (117 kg/sec)	3.5	11.94	53.8
HTGR	3.0	10.24	46.2
Water (155 kg/sec)	--	--	--
Total in	6.5	22.18	100.0
Product			
Hydrogen (247 m ³ /sec, 82.7 mol % H ₂) ^(a)	3.7	12.63	56.9
Tar (11.4 kg/sec)	0.4	1.36	6.1
Sulphur (3.5 kg/sec)	--	--	--
Tar gas/CO ₂ (183 kg/sec)	--	--	--
Subtotal	4.1	13.99	63.0
Export power	0.2	0.53	2.5
Waste heat	2.2	7.66	34.5
Total out	6.5	22.18	100.0

(a) At 273°K, 0.1013 MN/m².

TABLE 3-7
SCHEDULED OUTAGE TIME FOR A HELIUM OUTLET
TEMPERATURE OF 1366°K

Refueling periods per year	1
Refueling outage time (days)	22
Plant availability (%)	94
Percent of maximum load for availability (days)	85.1
Fuel capacity factor (%)	80.0

Reference 3-1 presents an extensive discussion on the design modifications necessary to reduce the peak fuel temperatures in the core below 1800°K (2780°F). The changes required in commercial HTGRs to make them suitable for this application are aimed at increasing the coolant gas outlet temperature of the core without increasing the fuel temperature and are listed below.

1. Changes could be made in the graphite block design to increase the number of fuel holes per block to 210. The disadvantages of this modification would be increases in core pressure drop, number of fuel rods to be produced, and number of holes to be drilled in the blocks. Consequently, fuel cycle costs would increase.
2. TRISO coatings could be required on all particles. The advantages of this modification would be reduced fuel rod shrinkage, which would reduce the temperature difference between the fuel rod and the graphite block, and reduction of the metallic fission product release. The disadvantage of this modification would be an increase in the amount of silicon, which is a neutron poison. This would also increase fuel cycle costs.
3. The C/Th ratio could be reduced from 240 to 200. This change would especially result in less power peaking, which would reduce the maximum fuel temperature. This reduction in the C/Th ratio would result in about a 2.4% reduction in radial power peaking factors.
4. Axial push-through fuel management with a 3-yr cycle could be accomplished. The principal advantages of this refueling scheme would be no age peaking, an approximately ideal power profile, a reduced fast fluence, a reduced core pressure drop, a smaller core, and a simplification of core design.

5. An improved fuel technology which would allow a peak fuel temperature of $\sim 1800^{\circ}\text{K}$ could be developed.

Very promising results have been found (Ref. 3-6) regarding the development of a suitable heat exchanger alloy which withstands corrosion in the fluidized bed, shows sufficient creep rupture strength, and allows tube forming. These alloys are still very basic and have to be proven on a pilot plant scale. Although the feasibility of the gas generator described in Ref. 3-3 has been proven, further pilot plant testing and improvement seem necessary. No basic problems are expected.

The process steps involved in product gas purification and shift conversion are based on a well proven technology, so it can be assumed that all problems have already been solved.

3.3.6. Status of Materials Technology

Materials technology has not kept up with the very fast development of fuel and core materials. Generally speaking, it can be said that a gap of 300° to 500°K (540° to 900°F) exists between the temperature capabilities of fuel and core materials and those of existing metals and alloys. Therefore, much attention has been and is still being given to

1. The effect of the helium environment on the structure of metals and alloys.
2. The influence of the fluidized bed environment (coal particles, steam, CO , CO_2 , H_2 , CH_4) on the heat exchanger materials.

Early work carried out by the Dragon research team (Ref. 3-5) has shown that nickel/chromium and iron/nickel/chromium alloys containing titanium and aluminium are severely attacked by impurities in helium at temperatures in excess of 1000°K (1340°F) whereas alloys containing

tungsten and niobium are not severely attacked. Furthermore, molybdenum and its alloys proved to be safe, but niobium alloys did not. The major problems appeared to be associated with metallurgical stability. Under the prevailing conditions, the primary coolant could be oxidizing and carburizing to certain elements.

The work carried out so far has indicated that existing alloys of appropriate strength are unlikely to give satisfactory performance in a nuclear installation at very high temperatures. However, enough basic information has been obtained to assess the feasibility of the very-high-temperature process heat reactor.

Experiments carried out in Germany (Ref. 3-6) have shown the influence of alloying elements on oxidation behavior and have indicated that suitable materials can be developed for heat exchangers.

The research and development program devoted to advanced HTGR fuel systems for process heat applications, core components, and materials research is described in Ref. 3-1.

3.3.7. Resource and Environmental Considerations

Coal gasification with nuclear heat undoubtedly has less environmental problems than conventional gasification processes. The emission of CO_2 per ton of gasified coal is remarkably reduced, and compared with direct combustion of solid and liquid fuels, sulfur emission is avoided. The utmost care has to be taken, however, to avoid tritium migration from the primary coolant loop via the secondary loop into the product gas. Another important advantage of the present process concerns its fresh water usage. The BFW requirements amount to 8.4 kg/sec per kg/sec of pure hydrogen compared with 25.6 kg/sec per kg/sec of pure hydrogen for the process described in Section 3.4. The same safety considerations which apply to the steam plant apply to the very-high-temperature gas-cooled reactor. Reference 3-1 gives an extensive description of the safety considerations for a very-high-temperature gas-cooled reactor.

3.4. HYDROGEN MANUFACTURE BY LURGI GASIFICATION OF OKLAHOMA COAL

3.4.1. Introduction

The Lurgi process for gasification of coal in a fixed bed reactor using steam and oxygen up to a pressure of $3 \times 10^6 \text{ N/m}^2$ (435 psia) was first used commercially in 1936. After World War II, the process was further developed and is now used on a large scale at several locations to produce synthesis gas and gas for public consumption.

In the present study, the Lurgi pressure gasification process is used to make a synthesis gas which is shift converted with steam to produce hydrogen. In general, gasification with oxygen and steam at an elevated pressure in fixed bed reactors has the following advantages (see Ref. 3-30):

1. There are high gasification rates per unit of reactor volume, which means a reduction in the percentage of heat lost from the reactor.
2. There is a minimum cost of compression because it is only necessary to compress the oxygen, which, in the process studied, amounted to 25.5% of the volume of the gas produced, to, $3 \times 10^6 \text{ N/m}^2$ (435 psi).
3. The gas purification and shift conversions can be carried out more economically because of the lower investment cost of smaller vessels.
4. Waste heat recovery and cooling of gas streams can be more easily performed at higher pressures.
5. Small-size fuel may be gasified with a minimum of carry-over because of the low gas velocities resulting from the higher gas density.

6. With countercurrent flow and gasification in the presence of excess carbon, hydrogenation reactions occur and the exothermal heat of these reactions is recovered, resulting in reduced oxygen requirements.

Owing to the higher gas residence time, less-reactive fuels and fuels rich in ash can be utilized.

The disadvantages of using pressure in fixed bed processes are

1. Pressure shifts the equilibrium of the hydrogenation reactions to increased yields of methane which, when hydrogen is the important product, must be removed from the product gas in some cases.
2. Some coals that are noncaking at atmospheric pressure develop caking tendencies when gasified under pressure.
3. Feeding of solid fuel into the pressure vessel is a somewhat complicated operation.

In general, hydrogen manufacturing processes involve three stages:

1. A main reactor stage, in which the hydrogen-containing material reacts to generate hydrogen:



2. A shift conversion stage, in which byproduct CO is converted to CO_2 and H_2 according to the water gas shift reaction:



3. A purification and compression stage, in which the H_2 product stream is further purified by means of cryogenic, absorption, adsorption, or diffusion techniques in order to meet the product specification.

In contrast with the other processes described in this study, the required heat for this process is not delivered by an HTGR, but is generated by burning part of the pretreated coal in the steam generator. The process feed is Oklahoma coal at 659 kg/sec (2613 tons/hr), and the product gas [$623 \text{ m}^3/\text{sec}$ (1901 MMscfd)] contains 81.0% H_2 at $10.3 \times 10^6 \text{ N/m}^2$ (1494 psi).

3.4.2. Process Description

A block flow diagram of the process is shown in Fig. 3-6 (see Ref. 3-31). The Oklahoma coal (see Table 3-8 for the analysis) is prepared by crushing to obtain particles with a diameter size of 3×10^{-3} to $12 \times 10^{-3} \text{ m}$ (9.8×10^{-3} to $39.2 \times 10^{-3} \text{ ft}$), screening the particles, and conveying them to a pretreatment section where mild oxidation with air takes place. The mild oxidation is intended to eliminate the tendency of the coal to agglomerate. The off-gas, which contains SO_2 from the pretreatment section, is sent to the sulfur recovery plant. Undersized coal from the screening operation is conveyed to a briquetting plant, where it is briquetted to a size acceptable for the gasification process. Pretreated coal (see Table 3-8 for analysis) at a rate of 423 kg/sec is conveyed to the gasifiers, where it is gasified at $31 \times 10^5 \text{ N/m}^2$ (450 psia), with oxygen and steam as the gasifying agents. The required gasification steam at 424 kg/sec is generated in the pretreatment section and the jacket and the waste heat boiler of each gasifier. Oxygen is manufactured in eight oxygen plants, each producing 23 kg/sec of 99.5%-pure oxygen (Ref. 3-32). Coal is fed to the gasifier (see Fig. 3-7) through a lock hopper at the top and successively passes the coal preheat zone and the reaction zone on its way down to the ash zone at the bottom of the gasifier.

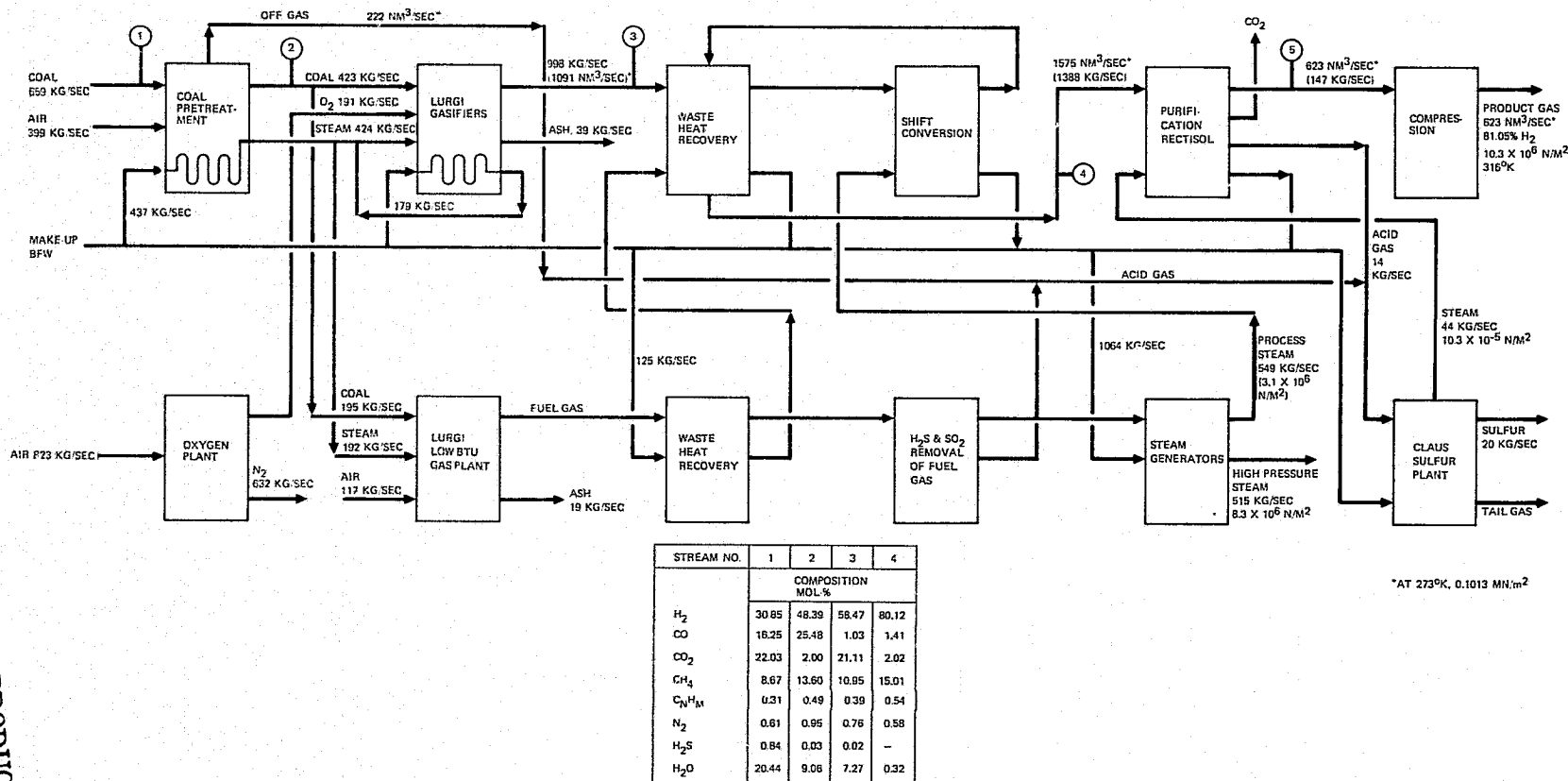
Fig. 3-6. H₂ production using Lurgi gasification

TABLE 3-8
HYDROGEN MANUFACTURE OF OKLAHOMA COAL
BY LURGI GASIFICATION

	Analysis of Oklahoma Coal (wt %)	
	Before Pretreatment (Stream No. 1)	After Pretreatment (Stream No. 2)
C	67.5	71.6
H	5.0	4.3
O	7.7	9.2
N	1.6	1.9
S	3.8	3.4
H ₂ O	6.4	--
Ash	<u>8.0</u>	<u>9.6</u>
	100.0	100.0
Gross heat of combustion (kcal/kg)	7070.7	7131.7

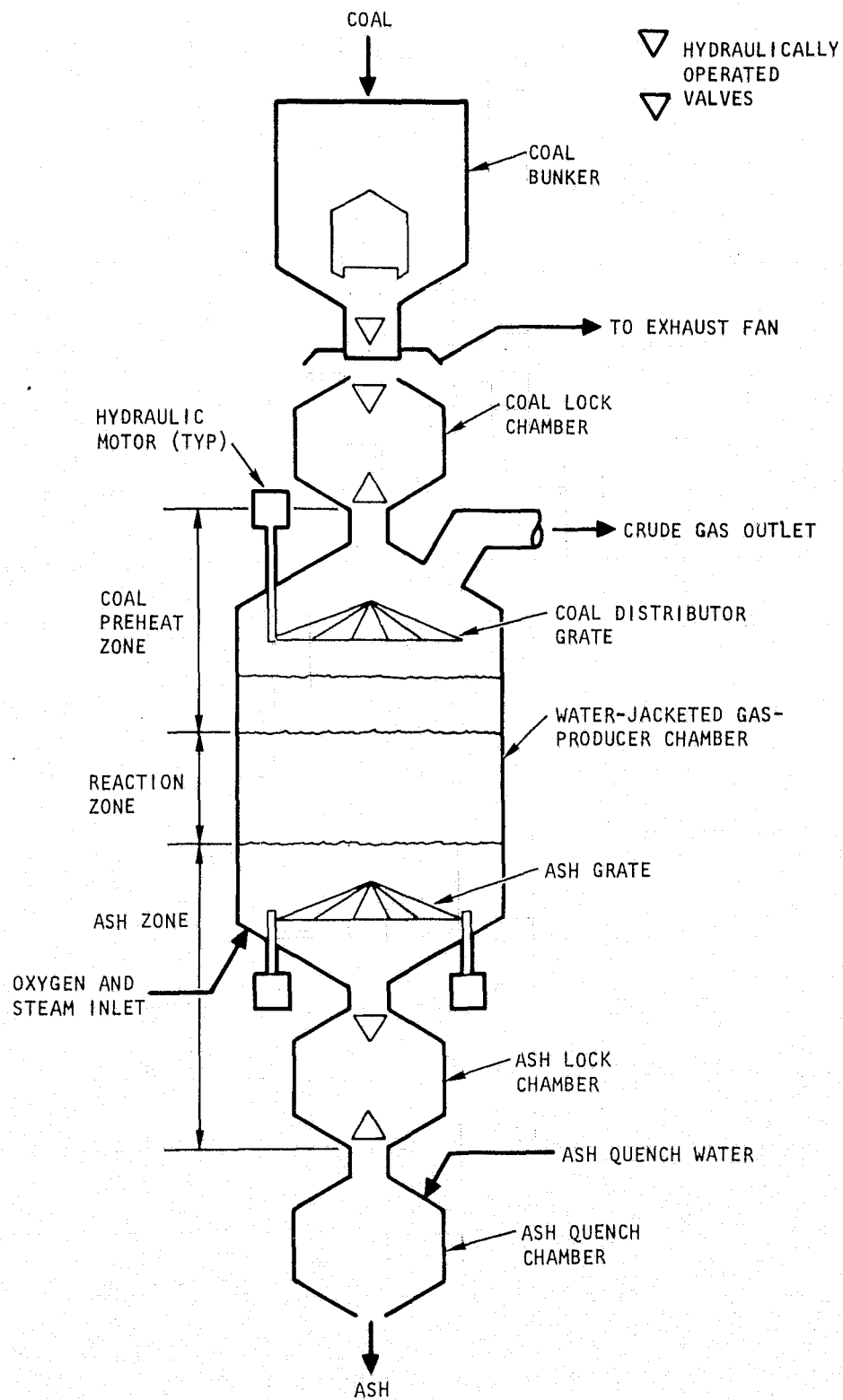
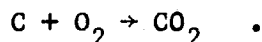
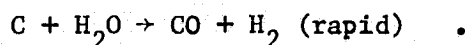
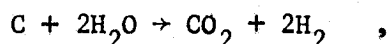
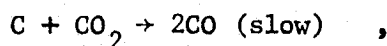
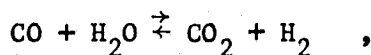


Fig. 3-7. Lurgi gasifier reactor

The steam/oxygen gasifying agent enters at the bottom of the gasifiers in the oxidation zone [peak temperatures are 1373°K (2012°F)], where the char is oxidized according to the equation



The gases leaving the oxidation zone enter the reduction zone, where the following reactions take place:



The movement of gases countercurrent to the flow of coal and ash causes the temperature of the gas stream to drop from the maximum temperature of 1400°K (2100°F) reached in the oxidation zone to about 820°K (1022°F) at the top. This is a distinct feature of the fixed bed process, accounting for the high thermal efficiency. About 95 gasifiers would be required to process 423 kg/sec of pretreated coal (Refs. 3-33, 3-34). The raw product gas leaves the gasifier at a temperature of 780°K (950°F) and enters the wash cooler, where impurities such as coal dust are removed with hot circulating water. Most of the tar containing the coal dust is recirculated to the gasifier (see Fig. 3-8). From the wash cooler, the product gas at 510°K (460°F) is heated in a waste heat recovery section to 615°K (645°F) combined with process steam at $31 \times 10^{-5} \text{ N/m}^2$ (450 psi) which has also been heated to 615°K (645°F), and fed to the shift reactors. The amount of steam added is sufficient to give a steam/CO molar ratio of 4.2/1. In the shift reactors, the steam reacts with CO to make more hydrogen according to the reaction



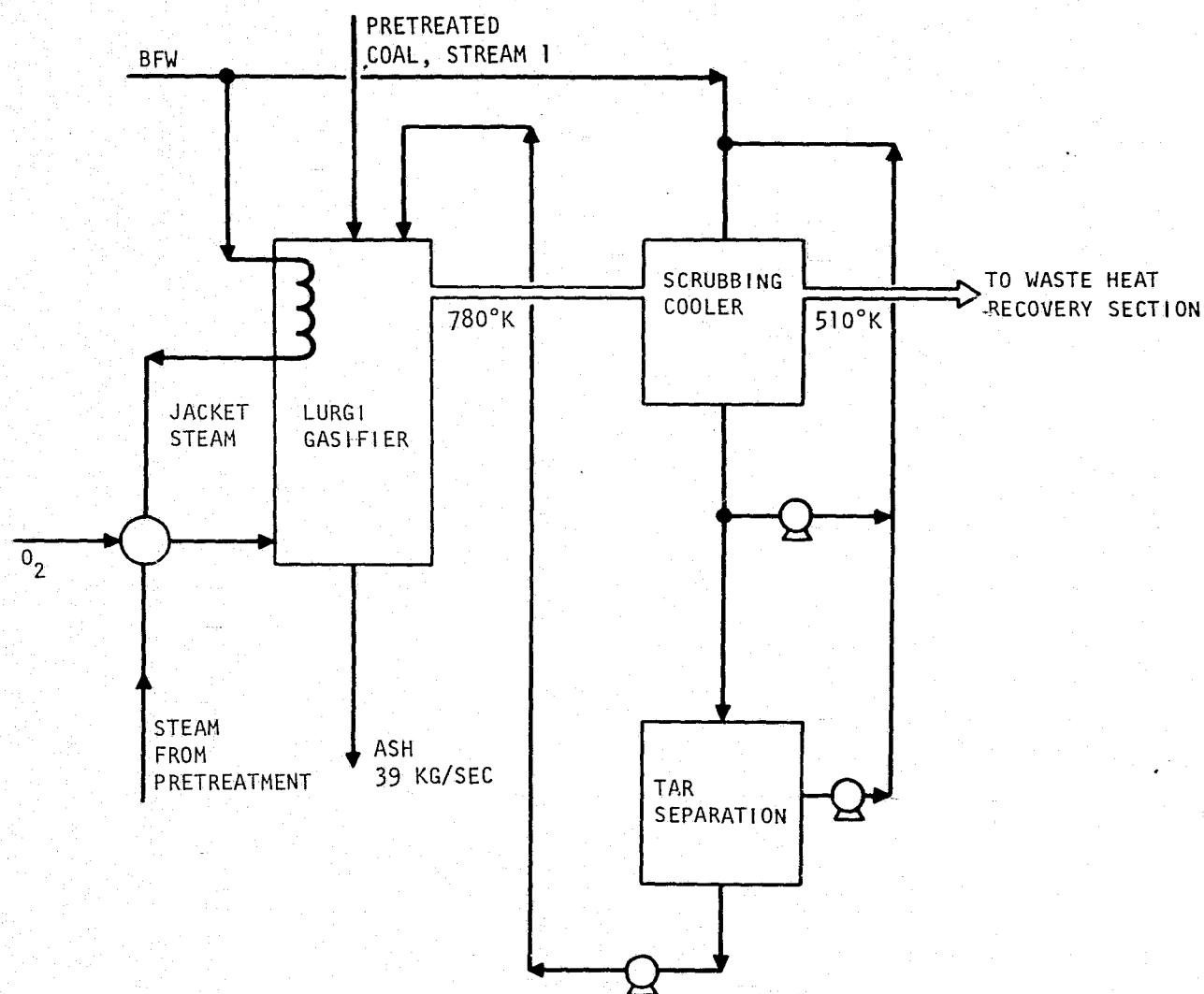


Fig. 3-8. Detail of Lurgi gasifier section

The reaction takes place over a specially developed, high-temperature shift catalyst, and 95% of the CO is converted. Twenty-two shift reactors are required (Ref. 3-35). The shift reaction is exothermic, and the product leaving the reactors at 753°K (~900°F) is cooled by exchanging heat with the gas feed stream to the shift reactors; this product is fed into the Rectisol gas purification units. In the first stage of the Rectisol units (Ref. 3-31), the gas is countercurrently washed with methanol at a temperature of about 200°K (-99°F), and the bulk of the CO₂, practically all the H₂S and hydrocarbons, and an appreciable amount of organic sulfur compounds are removed. The partially purified gas leaving the first stage flows to the second stage, where it is countercurrently washed with a stream of thoroughly stripped methanol which enters the column at about 210°K (-210°F). In this operation, most of the remaining CO₂ and practically all the residual organic sulfur compounds are removed from the gas. The solvent of the first stage is regenerated by two successive pressure reductions and flashing of the dissolved gases, and that of the second stage is stripped of acid gas by heating with indirect steam in a conventional stripping column, is cooled, and is recycled to the top of the second-stage contactor. The H₂S-rich gas is sent to Claus units for sulfur recovery. The analysis of the product H₂ stream is given in Table 3-9. Thirty-one Rectisol plants, each consisting of three purification units and one regeneration unit, are required to handle 1575 m³/sec (58,795 scf/sec) of gas coming from shift conversion. Finally, the H₂ product is compressed to the required pressure of 10.3×10^6 N/m² (1500 psi) and cooled to 316°K (110°F). The composition of the final product, listed in Table 3-9, is 81.0 mole % H₂ or 45.4 kg/sec of 100%-pure H₂.

Steam at 515 kg/sec and 8.3×10^6 N/m² (1200 psi) to drive the oxygen plant and product compressors and steam at 549 kg/sec and 3.1×10^6 N/m² (450 psi) for shift conversion are generated in the steam generation section, which consists of thirty 6-m-diameter (20-ft-diameter), Lurgi low-Btu fuel gas plants, each with its own waste heat recovery section in which about 55% of the steam required for gasification is raised. The outlet

TABLE 3-9
HYDROGEN MANUFACTURE OF OKLAHOMA COAL
BY LURGI GASIFICATION

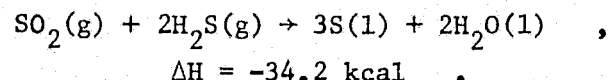
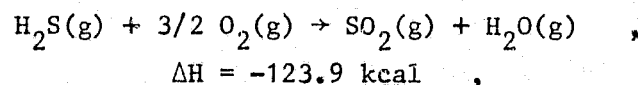
	Composition (mol %)		
	Stream No. (a)		
	3	4	5
	Washed Gas of the Gasifiers	After Shift Conversion	Product Gas
H ₂	30.85	32.05	81.05
CO	16.25	0.56	1.42
CO ₂	22.03	25.95	1.00
CH ₄	8.67	6.00	15.18
C _n H _m	0.31	0.21	0.28
N ₂	0.61	0.42	1.07
H ₂ S	0.84	0.58	--
H ₂ O	20.44	34.21	--

(a) Stream numbers refer to Fig. 3-6.

temperature of the waste heat boiler is $\sim 400^{\circ}\text{K}$ (260°F). The impurities of the product gas are removed by quenching and scrubbing with hot tar-containing water in the wash coolers. Most of the tar, containing coal dust and other impurities, is recirculated to the gasifiers. From the wash cooler, the producer gas is cooled in a series of heat exchangers and introduced in an absorber/stripper system which removes the H_2S . The sulfur is recovered in the Claus unit and the gas is fired in the steam generators, producing process steam and high-pressure steam.

In the present study, 30% of the coal fed to the pretreatment section is used to produce fuel gas for steam generation. An appreciable reduction of this percentage can be realized by recovery of methane-rich fuel gas from the product stream in, for example, a cryogenic separation unit. For an even bigger coal gasification plant, the obvious solution for process steam and high-pressure steam generation is a nuclear steam generating system, not only saving coal, but also reducing CO_2 emission and, depending on the coal price, eventually reducing the costs of gas production.

The sulfur recovery plant consists of three 7.7-kg/sec Claus units in which the produced H_2S is burned in specially constructed furnaces to give SO_2 , which is again converted into recovered sulfur in a second-stage converter. The reactions taking place are



The overall material balance is shown in Table 3-10. Six hundred fifty-nine kg/sec of Oklahoma coal is required to produce 147.2 kg/sec of product gas (81.0% H_2).

TABLE 3-10
SUMMARY OF OVERALL PERFORMANCE, PROCESS 4 (LURGI)

	Gross Heating Value		% of Total Input
	(GW)	(10 ⁹ Btu/hr)	
Input			
Coal (659 kg/sec)	19.5	66.5	100.0
Air (1339 kg/sec)	--	--	--
Water (747 kg/sec)	--	--	--
Total in	19.5	66.5	100.0
Product			
Hydrogen (623 m ³ /sec, 81 mol % H ₂) ^(a)	10.5	35.8	53.8
Subtotal	16.0	35.8	53.8
Waste heat	3.5	30.7	46.2
Total out	19.5	66.5	100.0

(a) At 273°K, 0.1013 MN/m².

Taking into account the necessary pretreatment of coal, the 8.3×10^6 N/m² (1200 psi) steam generation to drive the compressors in the oxygen plant and the product compressors, and the process steam for shift conversion, the overall thermal efficiency of the process is defined as

$$\eta = \frac{\text{Higher heating value of product (kcal/kg)}}{\text{Higher heating value of coal feed (kcal/kg)}} \times 100 \quad ,$$

which is equal to 53.8%.

3.4.3. Status of the Process Technology

The major limitation of the Lurgi process is that the size of fixed bed gasifiers is difficult to increase because of their mechanical complexity. This size is presently limited to an inner diameter of about 3.65 m (12 ft) (Ref. 3-31). In addition, fixed bed gasifiers are basically low-throughput devices. This means that for the present application, a large number of gasifiers are required, which in turn implies more complex and more costly piping than if fewer gasifiers were needed. The plant investment will consequently be high. A significant increase in the capacity of the gasifiers is possible by increasing the diameter. An enlargement to 6.5 m (21 ft) would result in an increase in coal capacity by a factor of 3.4 compared with that of the 3.65-m-diameter (12-ft-diameter) gasifiers. This implies that if 6.5-m-diameter (21-ft-diameter) gasifiers were available, the number of gasifiers could be reduced to about 28 to 30, which would result in a simpler operation without loss of flexibility.

Operating experience with large Lurgi plants with six 2.6-m-diameter (8.5-ft-diameter) generators for the production of pipeline gas has been obtained in Australia and Germany and with the SASOL-I plant, which consists of ten 3.7-m-diameter (12.1-ft-diameter) generators for Fisher-Tropsch synthesis, in South Africa. A plant of the size described in this

section will therefore be a significant increase. However, by using a modular design, piping problems can be limited.

According to the most recent information, the government of South Africa has decided to build a SASOL-II plant consisting of approximately 100 Lurgi gasifiers, i.e., a plant with a capacity ten times greater than that of SASOL-I. Although this plant will mainly focus on production of oil and gasoline from coal using the Fisher-Tropsch process, its overall size and complexity will be comparable to the hydrogen production plant described in this section (Ref. 3-36).

3.4.4. Duty Cycle

The large number of processing trains will guarantee a high overall plant availability and flexibility. Moreover, the Lurgi coal gasification route is based on a well proven process technology and years of operating experience. Special care has to be taken in the oxygen plants to remove traces of methane, ethane, and other hydrocarbons that are introduced in the airstream before entering the cold box of these plants. However, oxygen production is also based on a thoroughly proven process technology, and problems are not expected.

3.4.5. Resource and Environmental Considerations

The three most important environmental considerations are discussed below.

3.4.5.1. Water Usage and Preservation. Process water usage is equal to 747 kg/sec. No effort has been made in this assessment to reduce the total water usage for mining, processing, and ash quenching by treatment of the condensates. Only the condensates of the process steam turbines and those of the shift conversion reactors have been returned to the boiling feedwater line, and mechanical refrigeration has been chosen for the low temperatures needed in the Rectisol units.

According to a study made for the El Paso Natural Gas Company (Ref. 3-37), 1.5 kg of water are needed for each kilogram of coal to be gasified if every effort is to be made to conserve water usage and maximize recycle and reuse. This demand includes the water required by the coal mining operation, the gasification operation, and the auxiliary utility services. The ultimate disposition of the intake water is 10.2% used for process consumption, 69.6% returned to the atmosphere, 8.4% disposed to mine reclamation, and 11.8% for other uses.

It is obvious that for the plant described in this assessment, the utmost has to be done to reduce water usage, and the measures to be taken will be dependent on the location of the plant.

3.4.5.2. Stack Gas Emissions Containing SO₂, NO_x, and Particulates. Total stack gas emissions of sulfur will be about 0.023 kg/sec. Since there are no current federal, state, or local SO₂ emission standards for coal gasification plants, this figure has been compared with the emission standard for steam generators when coal is fired. According to this standard, the upper limit is 3.5 kg/sec of sulfur.

3.4.5.3. Solid Waste Disposal. Solid wastes from the process consisting of quenched ash from the gasifiers and steam generators can, in principle, be returned to the coal mining area for disposal. Plant location, with respect to the mining area, is very important because transport costs can significantly contribute to the operating costs of the plant.

3.5. HYDROGEN BY ELECTROLYSIS OF WATER

N76-15579

3.5.1. Introduction

The production of hydrogen and oxygen by electrolytic decomposition of water is a well established, developed technology. The electrical energy requirements for electrolysis are intrinsically large. As a consequence,

large-scale electrolytic hydrogen production today is limited to a few sites which provide low-cost hydroelectric power. Small-scale production is widespread where the high purity of electrolytic hydrogen or the ease of process control justifies premium costs.

The present assessment is concerned with the production of hydrogen by dedicated thermal-electric plants. Such plants are not normally characterized by very low power costs. As a consequence, the efficiency of electrical energy utilization assumes a greater importance relative to unit production rates. For assessment purposes, operating parameters are selected for efficient energy utilization.

Four overall systems are considered: two based on current electrolyzer technology using present efficiency values for electrical generation by fossil-fired and nuclear thermal stations and two using projected electrolyzer technology with advanced fossil and nuclear plants. These cases, in fact, resolve to two, since the efficiencies of the most efficient nuclear and fossil-fired plants are essentially equal.

In each case, a 3000-MW thermal input is assumed with a plant fully committed to hydrogen production. Hydrogen is delivered at 1.03×10^7 N/m² (1500 psia). Oxygen is vented to the atmosphere, and water is assumed to be available at a nominal energy cost.

3.5.2. Component Characteristics

3.5.2.1. Electrical Plant. The efficiency of current electrical plants is taken to be 39% for nuclear and fossil heat source cases. This value is consistent with recent fossil-fired facilities and the 1100-MW(e) HTGR. For advanced cases, a 50% efficiency is assumed. The technological basis is taken to be a gas-turbine-topped, fossil-fired, combined cycle and a gas turbine HTGR with a bottoming cycle.

3.5.2.2. Transformer/Rectifier System. For current technology, conversion to dc at appropriate voltage is assumed to be via rectification with a 95% efficiency. For the advanced systems, an efficiency of 97% will be assumed without defining the particular course of development. Direct generation as dc has been suggested, but this will entail substantial technological problems. For plants of the scale under consideration, the cell current to the electrolyzers is of the order of 10^9 A. Series connection above 500 V would entail significant safety and maintenance problems. At 500 V, copper buss bars on the order of 1 m (33 ft) in diameter would be required to maintain losses at 2% for transmission to 100 m (330 ft).

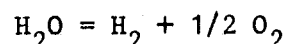
3.5.2.3. Compression. Product hydrogen is taken to be compressed in a four-stage compressor with inlet at and intercooling to 308°K (95°F). The machine efficiency is taken to be 85%.

Conventional and advanced electrolyzers can, in principle, deliver hydrogen at pressure. In this respect, electrolyzers can operate with an auxiliary compressor function close to unit efficiency. For the present assessment, the compressor alternative is elected because achieving the potential efficiency of pressure operation would entail efficient recovery of energy from the vented oxygen stream and, particularly for the advanced technology case, the excess energy associated with compression at elevated temperatures. The inefficiency of compression represents less than 2% of the primary energy input.

3.5.2.4. Electrolyzers. The current efficiency for electrolysis of water is essentially 100%. Consequently, the characteristic performance parameter for electrolysis is cell voltage. The theoretical minimum for cell voltage is given by the expression

$$V = \Delta G / 2F ,$$

where ΔG is the Gibbs free-energy change for the reaction



under cell conditions of temperature, pressure, and composition, and F is the Faraday constant. The ratio of theoretical voltage to operating voltage is frequently designated as the efficiency in electrochemical considerations. This electrochemical efficiency is related to the efficiency of the present appraisal as the Gibbs free energy of formation of water is related to the heat of formation of water.

If electrolysis were carried out at the theoretical potential, the energy produced would not be sufficient to supply the heat of decomposition of water, and heat addition would be required to maintain the cell at isothermal conditions. It is sometimes convenient to consider a voltage formally analogous to the theoretical minimum

$$V = \Delta H / 2F \quad .$$

This is the voltage for thermoneutral operation, i.e., isothermal, adiabatic electrolysis; ΔH is the heat of decomposition of water under cell conditions.

A variety of industrial-scale electrolyzers are available with the current technology (Ref. 3-7). These may be broadly classed as tank type and filter press. In the tank type, anodes and cathodes are alternated and separated by a diaphragm in a tank of electrolyte. The anodes and cathodes are electrically connected in parallel sets. The general arrangement is shown in Fig. 3-9. In filter press electrolyzers, the electrodes are bipolar, with one face serving as an anode and the other as a cathode. Diaphragms again define separate gas collection spaces. The general arrangement is shown in Fig. 3-9. By applying back pressure to the gas collection lines, filter press electrolyzers can be operated to produce product hydrogen and oxygen at pressure. Commercial units up to 3×10^6 N/m^2 (450 psi) are available.

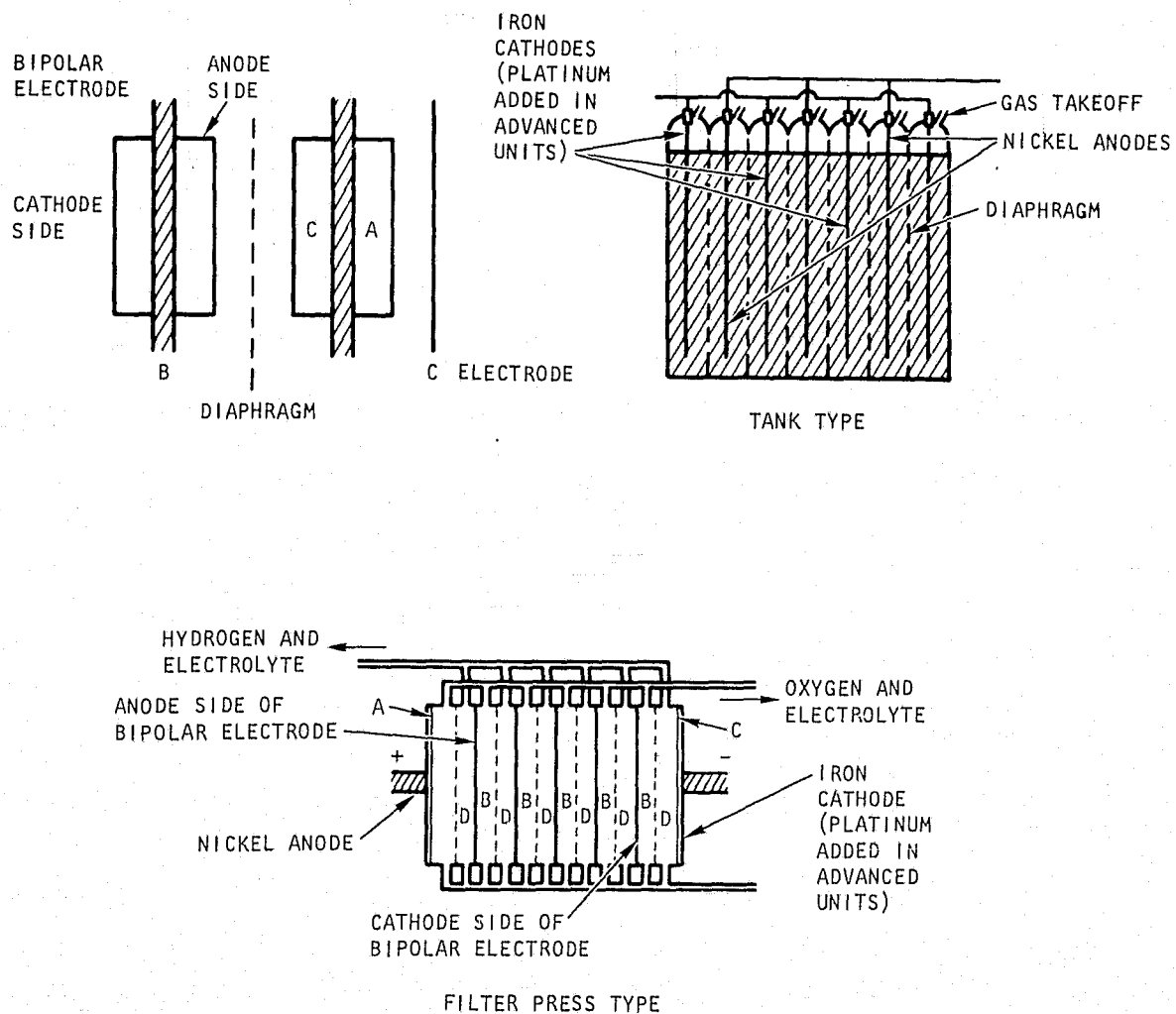


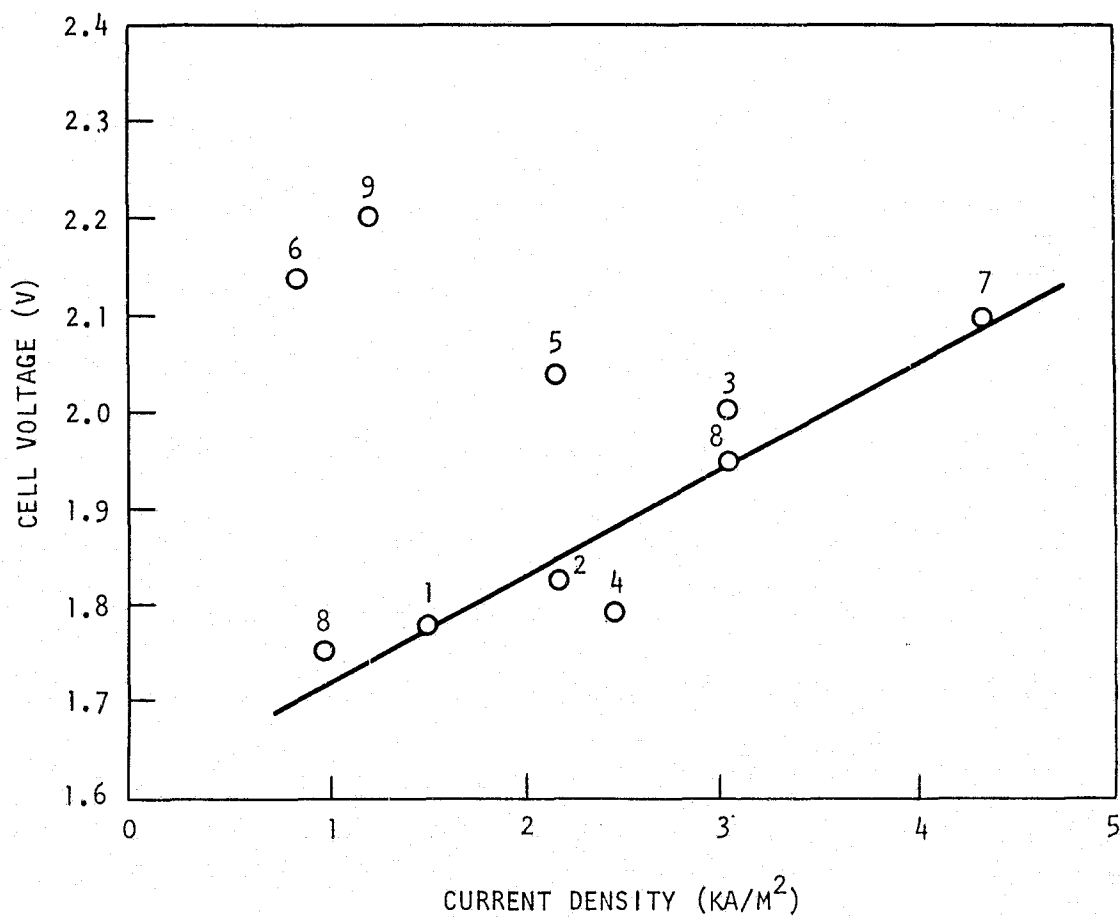
Fig. 3-9. Schematic of tank type and filter press type electrolyzer

In general, the electrolyte is a 25% to 30% aqueous solution of potassium hydroxide, and cell operating temperatures are on the order of 330° to 360°K (135° to 190°F).

Currently operating major electrolysis plants utilize cell voltages in the range of 1.8 to 2.2 V. As previously noted, these installations are located on sites of very low-cost power. For an energy-conservative appraisal, it must be assumed that lower current densities and cell voltages would be attained. An indication of plausible improvement can be obtained by examining the current voltage relation of existing plants as a group (Ref. 3-6), as shown in Fig. 3-10. In addition, DeNora (Ref. 3-8) indicated that cells operating at 1.61 V have been attained.

The cost of electricity of a dedicated plant considered in this study is expected to be about four times that of the present major hydroelectric sites. In a balanced system, this would call for a 50% reduction in current density. Combined with modest improvements in cell voltage, such as those implied by the DeNora claim, this leads to 1.65 V as a conservative cell voltage for a dedicated plant based on present technology. Although the conceptual plant discussion does not differentiate, the data of Fig. 3-10 suggest that a filter press type of cell would be required.

The extensive fuel cell development programs of the 1950s and 1960s brought forth basic changes in the approach to electrochemical systems. Some of these modified approaches have been applied to water electrolysis systems. Much of the impetus for this work has come from military and space requirements, but some of the resultant systems are relevant to potential large-scale production of hydrogen. Two advanced-technology systems which have been discussed are the Teledyne isotope system, which was derived from developments at Allis-Chalmers (Ref. 3-9), and the General Electric solid polymer electrolyte system (Refs. 3-10, 3-11). Of these two systems, the Teledyne isotope system has been developed further, and it has



TREND LINE EXCLUDES TANK-TYPE CELLS (5, 6, 9). CELL
 TYPES: 1 = HYDRO-PECHKRANZ; 2 = ZDANSKY-LONZA;
 3 = DENORA; 4 = BAMAG; 5 = STUART; 6 = TRAIL; 7 = EGGS;
 8 = DAMAG; 9 = KENT

Fig. 3-10. Cell voltage as a function of current density for existing plants

provided the basis for studies of hydrogen production by electrolysis. However, these studies have been constructed on an assumed, very low electrical energy cost.

The performance data presented by General Electric indicate a significant, but not overwhelming, improvement over the Teledyne isotope system on the basis of cell performance. General Electric workers have also speculated on the future technological development of their system.

Development projections suggest that in the year 2000, an improved cell based on an inorganic solid oxide conductor operating at 800°K (980°F) may produce hydrogen at an electrical energy consumption of 33 kWh/kg. For the present appraisal, a cell voltage of 1.277 V is used, corresponding to the thermoneutral voltage for an operating temperature of 800°K (980°F). This yields an electrical energy requirement of 34 kWh/kg.

The speculative nature of the technology forecasts should be noted. The potential utility of stabilized zirconia as a solid electrolyte was recognized at the turn of the century (Ref. 3-12), and its essential oxide-conducting character was established in 1943 (Ref. 3-13). Substantial efforts were made in the 1960s to develop a solid electrolyte fuel cell based on stabilized zirconia, and the technological difficulties of the extreme temperature required to attain adequate conductivity were recognized. Despite this, no alternative oxide conductor with the postulated conductivity at intermediate temperature has emerged even as a laboratory material. Projection into the higher temperature region implies a breakthrough rather than straightforward technological progress. Despite this problem, the postulated 1.277-V value remains plausible. When corrected for the thermal load required to bring the feed to cell conditions, the equivalent cell voltage is closer to 1.4 V. This is not much better than that reported for either the Teledyne isotope system or the General Electric solid polymer electrolyte cell at very low current densities.

An alternative path of development lies in the direction of cost reduction to permit utilization of low current densities. This path diverges from developments of the recent past which have emphasized system size and weight, with a consequent premium on high current densities. This emphasis has resulted in not only sophisticated fabrication but also use of significant amounts of precious metals as catalysts.

3.5.3. Results

Schematic flow sheets for current and advanced technology cases have been drawn, and heat and mass balances have been developed using the above assumed characteristics.

3.5.3.1. Current Technology. The flow sheet for the current-technology plant is presented in Fig. 3-11, and the results are summarized in Table 3-11. The entire thermal input is to the electrical plant, and a small portion of the electrical power, P_3 , is directed to the product hydrogen compressor. The balance is rectified and delivered to the electrolysis cells. Thermal balance requires that the electrical energy delivered to the cells in excess of the heat of formation of water be removed. In the conceptual flow sheet, this is achieved by vaporizing water into the product streams and condensing it as the gas streams are cooled. For an assumed cell voltage of 1.65 V, and using the properties of 30% potassium hydroxide, the resulting steady-state cell temperature is 359°K (187°F). This value is in the range reported for current electrolyzers. For pressurized cells, the temperature would rise, but can be controlled by taking a portion of the heat load, Q_3 , from cells or electrolyte recirculation lines.

For tank-type cells, the indicated electrolyte recirculation would be nominal. For filter press cells, recirculation volumes would be on the order of the product gas volumetric flow, operating primarily on a gas lift basis. This recirculation would be on a cell bank rather than a total recirculation basis. The overall efficiency of the system is 32%, based on the thermal input and the higher heating value of the product hydrogen.

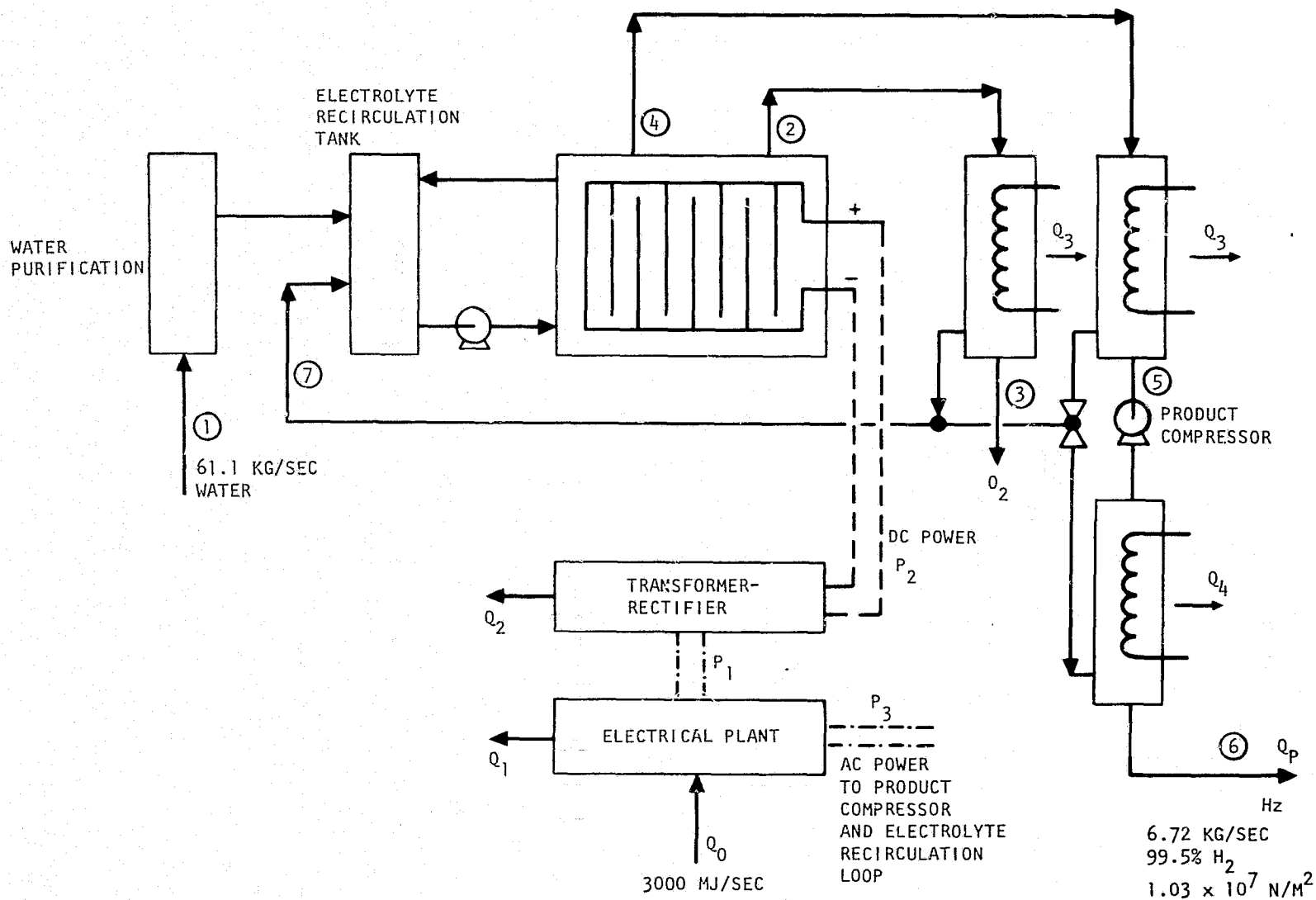


Fig. 3-11. Flow sheet, current-technology electrolysis plant

TABLE 3-11
MASS FLOWS AND ENERGY BALANCE, CURRENT-TECHNOLOGY PLANT(a)

Stream No.	Total Flow		Water		Hydrogen		Oxygen		Temperature	
	(kg/sec)	(lb/hr)	(kg/sec)	(lb/hr)	(kg/sec)	(lb/hr)	(kg/sec)	(lb/hr)	(°K)	(°F)
1	61.1	484×10^3	61.1	484×10^3					298	77
2	69.6	551×10^3	11.6	131×10^3			53.0	420×10^3	359	187
3	54.8	434×10^3	1.7	139×10^3			53.0	420×10^3	308	95
4	38.9	308×10^3	33.2	263×10^3	6.69	53×10^3			359	187
5	10.2	80×10^3	3.5	28×10^3	6.69	53×10^3			308	95
6	6.72	53×10^3	0.025	198	6.69	53×10^3			308	95
7	48.1	381×10^3	48.1	381×10^3					298	77

(a) All streams at nominal $1.01 \times 10^5 \text{ N/m}^2$ except 6, which is at $1.03 \times 10^7 \text{ N/m}^2$.

Energy Consumption/Rejection

$$Q_o = 3000 \text{ MJ/sec; } 10.2 \times 10^9 \text{ Btu/hr}$$

$$Q_1 = 1830 \text{ MJ/sec; } 6.2 \times 10^9 \text{ Btu/hr}$$

$$Q_2 = 56 \text{ MJ/sec; } 0.19 \times 10^9 \text{ Btu/hr}$$

$$Q_3 = 113 \text{ MJ/sec; } 0.39 \times 10^9 \text{ Btu/hr}$$

$$Q_4 = 22 \text{ MJ/sec; } 0.075 \times 10^9 \text{ Btu/hr}$$

$$P_1 = 1110 \text{ MW}$$

$$P_2 = 1054 \text{ MW}$$

$$P_3 = 60 \text{ MW}$$

$$Q_p = 948 \text{ MJ/sec; (HHV) } 3.24 \times 10^9 \text{ Btu/hr}$$

$$Q_p/Q_o = 0.316$$

3.5.3.2. Advanced Technology. The schematic flow sheet for the conceptual advanced system is presented in Fig. 3-12, and the results are summarized in Table 3-12. In this case, a small portion of the thermal input, Q'_0 , is required to bring the water feed stream to cell conditions. In an optimized system, this heat might be drawn from a lower temperature in the thermal cycle, with a reduction in the net loss of electrical plant output. Eighty percent of the sensible heat of the product gases, Q_3 , is taken to be recovered by exchange with the feed stream.

Because of the conjectural nature of the advanced technology system, any particular value for the recycle flow of water would be arbitrary. An equally arbitrary value of zero was taken for simplicity: i.e., it is assumed that the feed is completely electrolyzed. This is forbidden thermodynamically, since the implied cell voltage in the terminal stage is infinite; it is also forbidden by mass transport polarization.

With solid oxide conductors, a 90% conversion would be plausible. With solid polymer electrolytes which require water and can be dehydrated, gaseous feed operation might be degraded at lower conversions.

If 80% heat recovery from the recycle stream is assumed, a system operating at 90% conversion would have a production rate of 99.5% of that developed in the present analysis. Even with a conversion of only 50%, the production rate would only be reduced by about 4.4%. The overall efficiency of the system is 49% based on the total thermal input, $Q_0 + Q'_0$, and the higher heating value of the product hydrogen.

3.5.4. Resource Impact

At the projected efficiencies, electrolytic production of hydrogen with its subsequent use as fuel would be more energy consumptive than electrical heating, not including electrical transmission losses, which can be significant. If a nuclear prime source of energy is assumed,

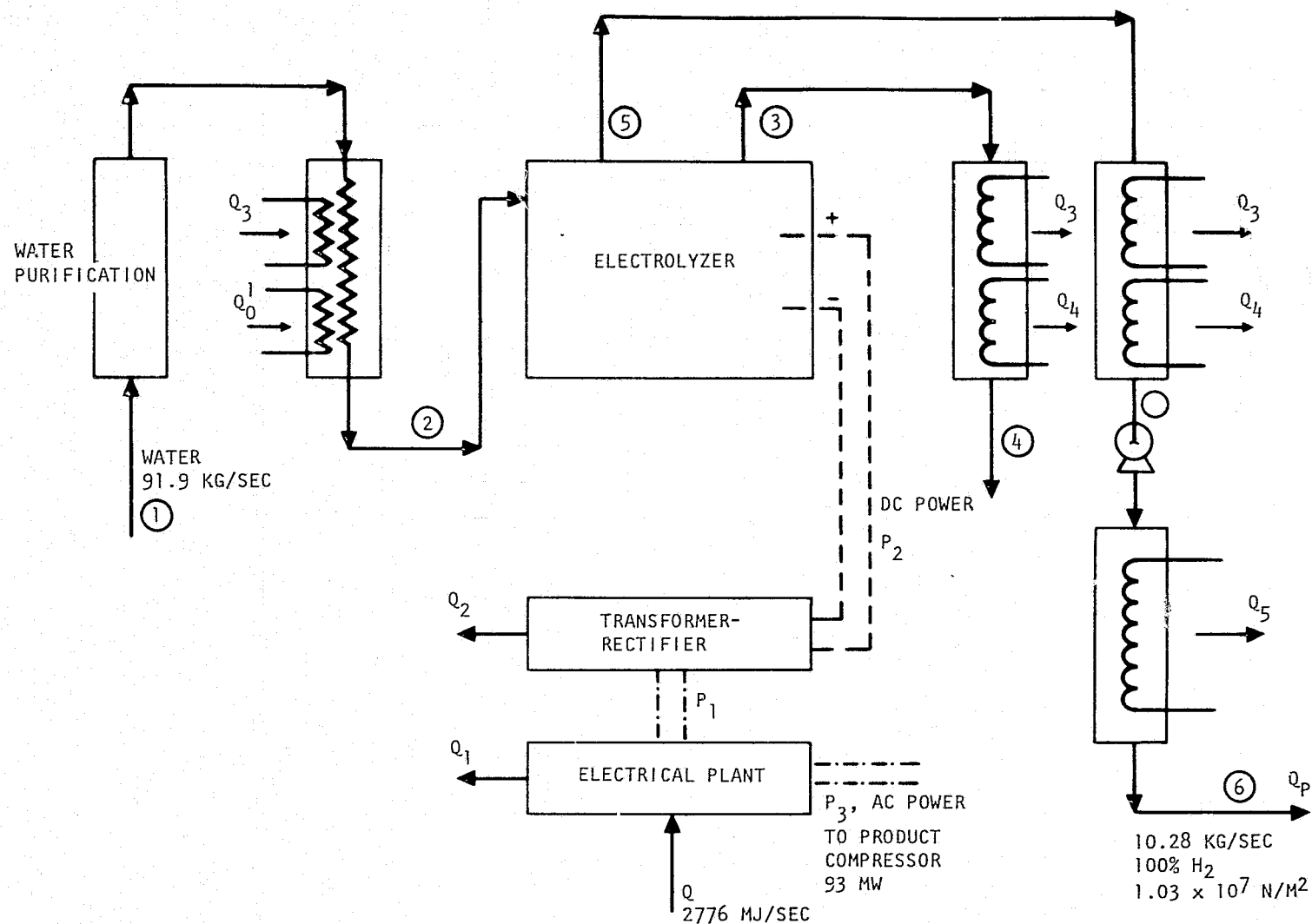


Fig. 3-12. Flow sheet, advanced-technology electrolysis

TABLE 3-12
MASS FLOW AND ENERGY BALANCE, ADVANCED-TECHNOLOGY PLANT^(a)

Stream No.	Total Flow		Water		Hydrogen		Oxygen,		Temperature	
	(kg/sec)	(lb/hr)	(kg/sec)	(lb/hr)	(kg/sec)	(lb/hr)	(kg/sec)	(lb/hr)	(°K)	(°F)
1	91.9	727×10^3	91.9	727×10^3					298	77
2	91.9	727×10^3	91.9	727×10^3					800	980
3	81.6	646×10^3					81.6	646×10^3	800	980
4	81.6	646×10^3					81.6	646×10^3	303	95
5	10.28	81.4×10^3			10.28	81.4×10^3			800	980
6	10.28	81.4×10^3			10.28	81.4×10^3			298	77

(a) All streams at nominal $1.01 \times 10^5 \text{ N/m}^2$ except 6, which is at $1.03 \times 10^7 \text{ N/m}^2$.

Energy Consumption/Rejection

$$Q_o = 2776 \text{ MJ/sec; } 9.5 \times 10^9 \text{ Btu/hr}$$

$$Q_o' = 224 \text{ MJ/sec; } 0.76 \times 10^9 \text{ Btu/hr}$$

$$Q_o + Q_o' = 3000 \text{ MJ/sec; } 10.2 \times 10^9 \text{ Btu/hr}$$

$$Q_1 = 1388 \text{ MJ/sec; } 4.74 \times 10^9 \text{ Btu/hr}$$

$$Q_2 = 39 \text{ MJ/sec; } 0.13 \times 10^9 \text{ Btu/hr}$$

$$Q_3 = 92 \text{ MJ/sec; } 0.31 \times 10^9 \text{ Btu/hr}$$

$$Q_4 = 23 \text{ MJ/sec; } 0.08 \times 10^9 \text{ Btu/hr}$$

$$Q_5 = 34 \text{ MJ/sec; } 0.12 \times 10^9 \text{ Btu/hr}$$

$$P_1 = 1388 \text{ MW}$$

$$P_2 = 1256 \text{ MW}$$

$$P_3 = 93 \text{ MW}$$

$$Q_p = 1457 \text{ MJ/sec; (HHV) } 4.97 \times 10^9 \text{ Btu/hr}$$

$$Q_p/Q_o + Q_o' = 0.486$$

electrolysis does provide a transportable fuel supply which is not fossil fuel dependent. Indeed, electrolysis is the only proven technology that achieves that end. This course of development would, however, further increase the demand on uranium supplies.

If electrolytic hydrogen production is to have a major impact on the economy, elimination of precious metals in the development of advanced electrolyzers will be necessary or at least highly desirable. Twenty plants the size of the advanced technology plant would produce 6.5×10^9 kg/yr of hydrogen. This is near the lower end of the range estimated for U.S. ammonia production in the year 2000.

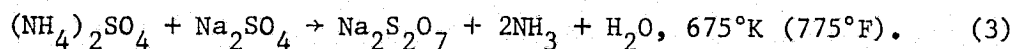
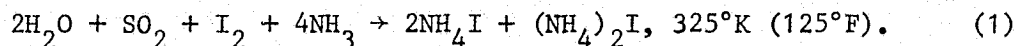
Assuming a current density of 10^4 A/m² (1000 mA/cm²) and a platinum loading of 1 mg/cm², these plants would require approximately 20×10^3 kg of platinum. This was approximately the total purchase of platinum by consuming industries in the U.S. in 1972 (Ref. 3-14). Such a complex of plants could undoubtedly come on-stream over a period of years, but the impact on platinum resources would still be significant.

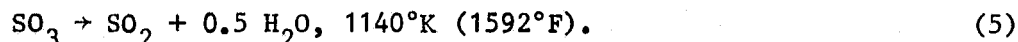
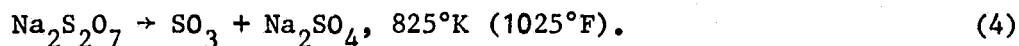
3.6. ISPRA MARK-10 WATER SPLITTING PROCESS

N76-15580

3.6.1. Summary

A process flow diagram, a material balance, and an energy balance were developed for the Ispra Mark-10 water splitting thermochemical reaction cycle. The principal reactions which constitute this cycle are





Reaction 1 proceeds to completion in aqueous solution. The salts are dried, and NH_4I is removed by ethanol extraction. Reaction 2 reaches equilibrium at about 25% decomposition of HI . The extent of decomposition of ammonia is kinetically controlled and is a significant consideration which is not yet well defined. Reaction 3 occurs in a mixture of salt solids and proceeds to about 90% completion under practical conditions. Reaction 4 proceeds to near completion, and reaction 5 requires a catalyst and approaches equilibrium at about 80% completion.

The process converts 23% of the heat from a 3000-MW(t) HTGR into heat of decomposition of water and work of compression of elemental hydrogen at 300°K (80°F) and $2.1 \times 10^5 \text{ N/m}^2$ (30 psi) and elemental oxygen at 300°K (80°F) and $5 \times 10^6 \text{ N/m}^2$ (725 psi). The product hydrogen output rate is 4.7 kg/sec at 99.99% purity. The process vent off-gas waste stream contains waste hydrogen at an output rate of 0.024 kg/sec.

Adherence to process conditions which do not radically differ from those used experimentally by Ispra leads to a large inventory of iodine, as the element and as iodide, in the system. About 10% of the annual U.S. production of iodine is required as inventory to perform only reaction 2 at the process design rate. Another consequence of adhering to the processing conditions used at Ispra is the occurrence of large evaporative heat loads associated with producing the dry, separated NH_4I and $(\text{NH}_4)_2\text{SO}_4$ salts. Foreseeable experimental developments related to these problems should provide significant improvements in overall heat utilization efficiency, iodine inventory reduction, and capital-intensive vapor compression heat recovery requirements for the liquid evaporation steps. However, the heat load and handling difficulties associated with recycling a large amount of NH_4I for reaction 2 appear to be an inherent and significant impediment to

achieving high heat conversion (approaching or exceeding 40%) by the Ispra Mark-10 process.

Normal process wastes are cooling water not exceeding 325°K (125°F), oxygen product with not more than 50 ppm NH₃, and 4 ppm (NH₄)₂SO₄, and process vent off-gas waste stream. The process vent off-gas waste stream consists of 87 wt % O₂, 10 wt % H₂O vapor, and 3 wt % H₂. If undiluted, the total flow rate is 0.89 kg/sec. This mixture is potentially explosive and requires dilution or other safety provision. Under credible abnormal operating conditions, nitrogen tri-iodide might be generated. Special control and monitoring precautions are needed to preclude the formation of this explosive compound.

3.6.2. Introduction

A thermochemical water splitting process, the Ispra Mark-10 chemical reaction cycle, was chosen for examining the possibility of using water to produce hydrogen on a large scale for fuel and major industrial chemical uses. The assumed energy source for the process is an HTGR (helium cooled).

The Ispra Mark-10 thermochemical water splitting cycle involves reactions 1 through 5. Summation of these reactions leads to cancellation of all constituents except water, hydrogen, and oxygen:



If this overall net reaction occurs at 298.16°K (77°F) and 1.013×10^5 N/m² (14.7 psi) with the separate pure materials in their normal phase state at these conditions, 1.419×10^8 J of heat per kilogram of hydrogen must be supplied to the reaction.

Associated with this amount of heat (enthalpy) is an amount of work which is the Gibbs free energy, which is 1.177×10^8 J/kg H₂. The free

energy is that portion of the total energy (enthalpy) which can be converted to pure work under certain thermodynamically ideal conditions. It is almost 83% of the total heat energy (enthalpy) in the combustion of hydrogen to water.

Thermochemical water splitting involves only heat input, in an overall conceptual sense. The maximum thermodynamically possible yield of hydrogen from a given quantity of heat is related through the Carnot efficiency of a thermodynamically ideal engine to the temperature at which that heat is supplied:

$$\eta = W/Q_1 = (T_1 - T_2)/T_1 ,$$

where η = Carnot (ideal) efficiency,

W = maximum thermodynamically possible work; in the case of H_2 combustion at constant pressure and temperature, the same as the free energy,

Q_1 = amount of heat input to engine or process,

T_1 = absolute temperature of heat input,

T_2 = absolute temperature of waste heat removed from engine or process; in this case, the same as the process cooling water temperature.

The application of the above formula to coupling the water splitting process with the reactor heat source and the environment (in effect, the process cooling water) is complicated by three circumstances:

1. Heat input occurs over a range of temperatures.
2. Heat rejection into the cooling water occurs through several different temperature drops which drive the heat transfer from the process equipment internals into the cooling water.

3. Energy conversion inefficiencies occur whenever process material is transferred from a lower pressure to a higher pressure, or vice versa.

The effects of varying the heat input and output temperatures or temperature drops are discussed further in Sections 3.6.7 and 3.6.8. However, other process inefficiencies are scattered throughout the process. Estimating many of these inefficiencies involves lengthy iterative calculations, and in some cases it is neither appropriate nor feasible to generate a quantitative estimate of the efficiency until one is well into detailed specific equipment design.

The ability to convert heat into work (or free energy) is diminished in actual practice by the occurrence of temperature drops to drive heat transfer, pump and power generator equipment inefficiencies, and friction whenever anything moves in contact with something else. Some of the more readily calculable inefficiencies in the process have been partially accounted for, but some significant efficiency losses were omitted.

There are a sufficient number of major processing steps (22) for it to be convenient to think of the system as consisting of five subsystems:

1. Water splitting reaction subsystem.
2. Hydrogen production subsystem.
3. Hydrogen purification subsystem.
4. Oxygen production subsystem.
5. Oxygen purification subsystem.

Figure 3-13 illustrates the principal material transfer relationships among these subsystems.

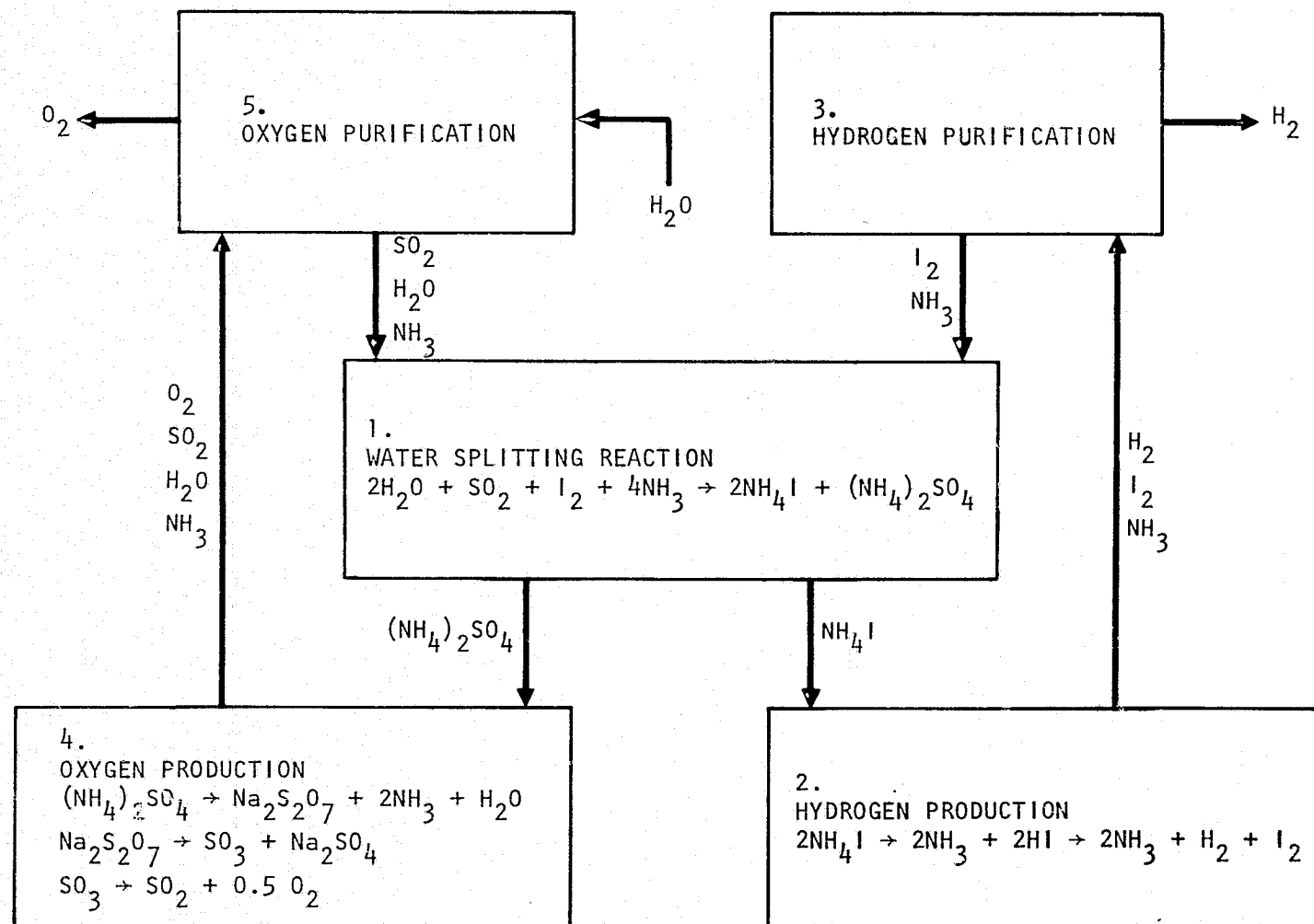


Fig. 3-13. ISPRa MKIO major processing steps

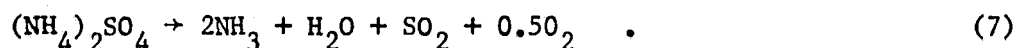
In the water splitting reaction subsystem, one of the two water molecules shown in reaction 1 is split. Its hydrogen becomes part of the NH_4I product, and its oxygen becomes part of the $(\text{NH}_4)_2\text{SO}_4$ product. The other water molecule is wholly incorporated into the $(\text{NH}_4)_2\text{SO}_4$. Separation of NH_4I and $(\text{NH}_4)_2\text{SO}_4$ is also done in the water splitting reaction subsystem. The NH_4I is transferred to the hydrogen production subsystem, and the $(\text{NH}_4)_2\text{SO}_4$ to the oxygen production subsystem.

In the hydrogen production subsystem, hydrogen is evolved from NH_4I by heating the salt to 900°K (1160°F). The principal decomposition products which result are H_2 , I_2 , and NH_3 . These decomposition products are transferred to the hydrogen purification subsystem, where hydrogen is separated from iodine and ammonia. Iodine and ammonia are recycled to the water splitting reaction subsystem, and purified hydrogen is the output product of the process.

In the oxygen production subsystem, the sulfate portion of ammonium sulfate is separated from the ammonia portion by reacting ammonium sulfate with sodium sulfate (reaction 3). Additionally, the water molecule which is not split (the second of the two water molecules indicated in reaction 1) is evolved by reaction 3.

Reaction 4 is then carried out in order to separate oxygen (which is split from the first of the H_2O molecules in reaction 1) from the Na_2SO_4 needed for recycle to reaction 3. Although individual atoms cannot actually be identified, it is conceptually convenient to think of the oxygen split from the water molecule in reaction 1 as being in the SO_3 evolved by reaction 4.

The oxygen (which is split from the first of the two water molecules in reaction 1) is then produced as free gaseous O_2 by decomposing SO_3 from reaction 4 with heat at 1140°K (1592°F). This final decomposition reaction is symbolized by reaction 5. The overall effect of reactions 1, 4, and 5 is to decompose $(\text{NH}_4)_2\text{SO}_4$ into NH_3 , H_2O , SO_2 , and O_2 :



The two decomposed reaction mixtures (most of the NH_3 and the H_2O are evolved in a stream which is physically separate from the $\text{SO}_2 + \text{O}_2$ stream) are transferred to the oxygen purification subsystem, where the oxygen is purified first by partial condensation of SO_2 and then by scrubbing with water. The condensed SO_2 and the spent scrub liquor are returned to the water splitting reaction subsystem, where they are needed as reactants for reaction 1. The purified oxygen is a product of the process.

The foregoing summary description of the Ispra Mark-10 water splitting process omits many important details in order to concisely describe the overall nature of the process functional requirements. Details needed for estimating the practical feasibility of fulfilling these functional requirements are described in Sections 3.6.3 and 3.6.4.

3.6.3. Assumptions and Calculation Bases

Heat was assumed to be supplied by a 3000-MW(t) HTGR. The helium outlet temperature from and the inlet temperature to the reactor were assumed to be 1255.4°K (1800°F) and 773.2°K (932°F), respectively. The reactor was assumed to have a duty cycle of 80% at full power. Low-power operation was not considered.

Except for the waste heat rejection heat exchangers having 300° or 325°K (80° or 125°F) process steam temperatures, temperature drops for heat transfer were taken to be about 50°K (90°F). Cooling water at 290°K (62°F) was assumed to be available in any amount desired for the waste heat exchangers. Waste heat disposal is discussed further in Section 3.6.7.

Power conversion efficiency in compressors, pumps, and power generation devices was assumed to be 80% of the thermodynamic ideal efficiency. No allowances were made for power consumption in fluid flow

frictional pressure drops, general plant utility requirements, or direct heat loss from the surfaces of hot process equipment.

Process conditions specified in information available from Ispra were used in this work unless there was a compelling reason for departure from these conditions. The main departures involved increased process pressures in cases where the implied process pressure of the Ispra data was $1.013 \times 10^5 \text{ N/m}^2$ (14.7 psi). Pressure increases are needed in some parts of the process to permit reasonably complete material separation by standard engineering methods such as condensation and distillation. In other parts of the process, pressure increases are needed for vapor compression to make possible recovery of the heat of condensation at temperatures high enough for reuse within the process. The highest pressure required in the process is $2.5 \times 10^7 \text{ N/m}^2$ (3600 psi) for resynthesizing cracked ammonia by a Haber cycle type of process.

Some general constraints which were placed on the form of the Ispra Mark-10 flow sheet developed in this report are

1. To make the process self-contained.
2. To adhere as nearly as possible only to conditions stated or implied by Ispra.
3. To specify operations and conditions having reasonable prospects of engineering feasibility.

In some cases, these goals conflicted, and it was necessary to draw compromises such that no single goal was as well optimized as would have been possible by disregarding the others.

3.6.4. Subsystem Functional Descriptions

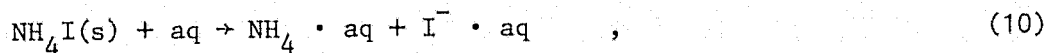
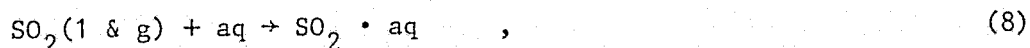
3.6.4.1. Water Splitting Reaction Subsystem. The water splitting reaction subsystem consists of four principal process steps:

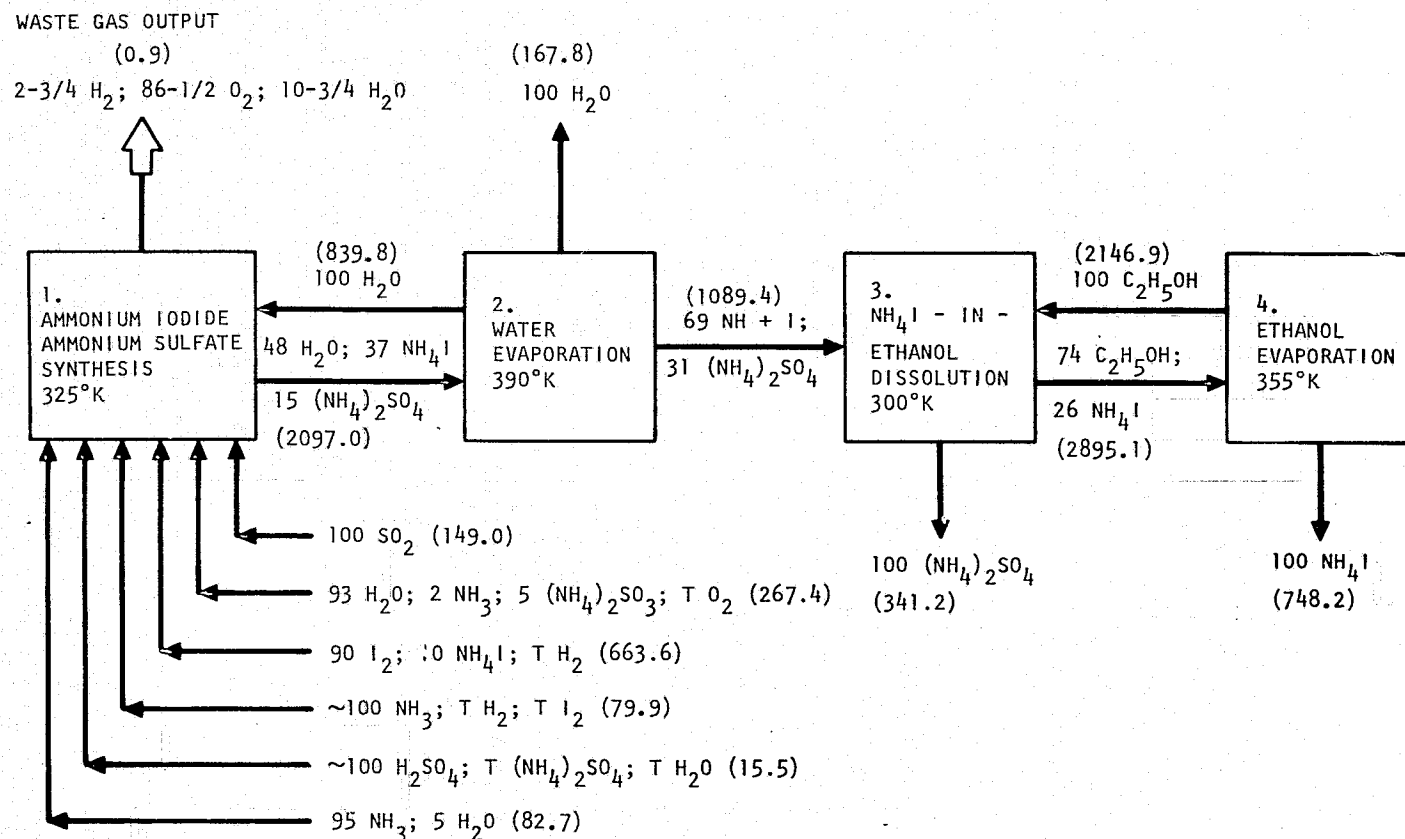
1. Ammonium sulfate -- ammonium iodide synthesis.
2. Water evaporation.
3. Ammonium iodide dissolution in ethanol.
4. Ethanol evaporation.

Figure 3-13 illustrates the material transfer relationships among these principal process steps.

Two constituents, NH_4I and H_2SO_4 , are shown as inputs on Fig. 3-14 but not on Fig. 3-13. Ammonium iodide acts as a catalyst for the reaction by solubilizing the elemental iodine in aqueous solution. H_2SO_4 recycle occurs because reaction 3 in subsystem 4 is actually a two-step reaction, the first step of which evolves water but does not proceed to completion. There would be little or no H_2SO_4 recycle if this first reaction step went to completion before the second reaction step occurred.

The reactions which occur in ammonium iodide - ammonium sulfate synthesis are

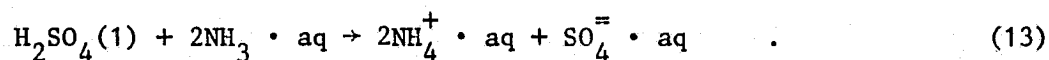
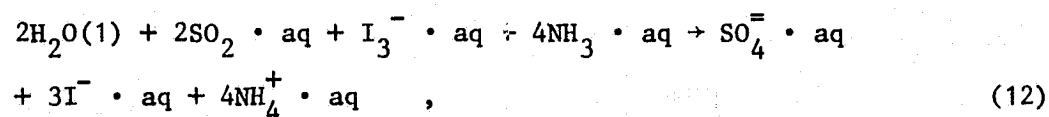




NOTES:

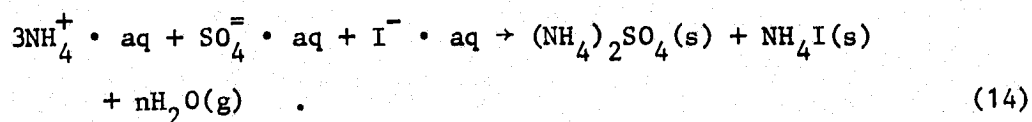
SUBSYSTEM PRESSURE IS NOMINALLY $1.013 \times 10^5 \text{ N/M}^2$ THROUGHOUT.
NUMBERS ASSOCIATED WITH STREAM COMPONENTS ARE APPROXIMATE WT % COMPOSITIONS..
NUMBERS IN PARENTHESES ARE APPROXIMATE TOTAL MASS FLOW RATES IN KG/SEC.
T = TRACE.

Fig. 3-14. Water splitting reaction subsystem

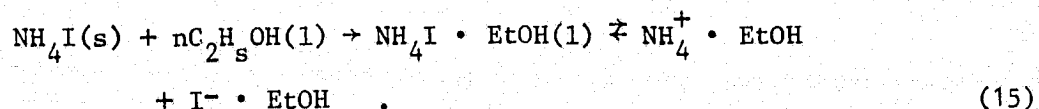


Nominal reaction temperature and pressure are 325°K (125°F) and $1.013 \times 10^6 \text{ N/m}^2$ (14.7 psi), although any convenient temperature-pressure combination which allows liquid solution to be the stable phase should be acceptable.

The reaction which occurs with evaporation of water from the solution made in the preceeding step is

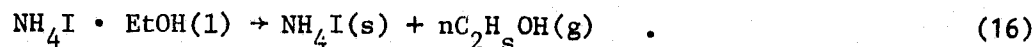


Since $(\text{NH}_4)_2\text{SO}_4$ is insoluble in ethanol and NH_4I is highly soluble, the salts are separated by an ethanol extraction step which may be represented as follows:



It is not known to what extent, if any, NH_4I ionizes in ethanol.

After separating the $\text{NH}_4\text{I} \cdot \text{EtOH}$ solution from the $(\text{NH}_4)_2\text{SO}_4$ solids, NH_4I is again recovered as a solid by evaporation:



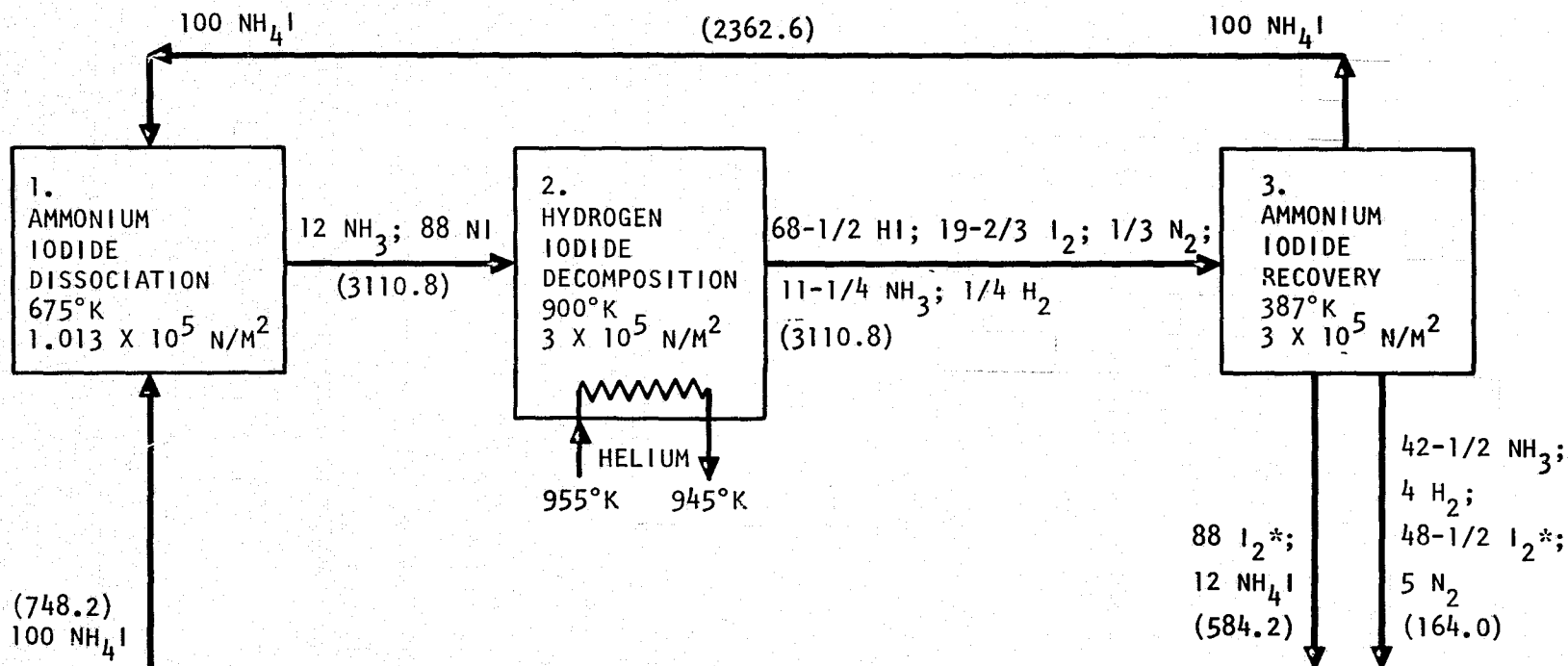
The evaporation temperatures in the evaporation steps are specified to be somewhat higher than the boiling points of the pure solvents [373.2°K (212°F) for water and 351.6°K (173°F) for ethanol]. Solutions boil at somewhat higher temperatures than pure solvents, but the main reason for specifying somewhat elevated temperatures is to desiccate to some extent the solids of the adsorbed solvent after they achieve superficial apparent dryness. It is especially desirable that water be kept out of the ethanol, since water is a solvent for $(\text{NH}_4)_2\text{SO}_4$ and would lead to the appearance of sulfate in the NH_4I salt. The effects of imperfect separations are given in Section 3.6.5.2.

3.6.4.2. Hydrogen Production Subsystem. The hydrogen production subsystem consists of three principal processing steps:

1. Ammonium iodide dissociation.
2. Hydrogen iodide decomposition.
3. Ammonium iodide recovery.

Figure 3-15 illustrates the material transfer relationships among these principal process steps.

Dry NH_4I solids are received from the water splitting reaction subsystem and are totally vaporized (dissociated) before adiabatic compression to 900°K (1160°F) and $3 \times 10^5 \text{ N/m}^2$ (44 psia). The increased pressure in the decomposition step improves the kinetics of HI decomposition, but the main reason for compressing the reaction mixture is to make the heat of condensation of undecomposed NH_4I present in the effluent from the HI decomposer as $\text{NH}_3 + \text{HI}$ available for recovery and



NOTES:

NUMBERS ASSOCIATED WITH STREAM COMPONENTS ARE APPROXIMATE WT % COMPOSITIONS.

NUMBERS IN PARENTHESES ARE APPROXIMATE TOTAL MASS FLOW RATES IN KG/SEC.

*~87% OF TOTAL I_2 IN BOTH EFFLUENTS.**~13% OF TOTAL I_2 IN BOTH EFFLUENTS.

Fig. 3-15. Hydrogen production section

reuse in the NH_4I dissociation step, which occurs at a pressure of $1.013 \times 10^5 \text{ N/m}^2$ (14.7 psia).

The reaction which occurs in the NH_4I dissociation step is assumed to be:



Data in Ref. 3-15 indicate the possibility that an equilibrium exists in the gas phase among the following species:



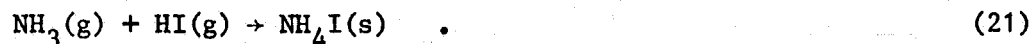
Complete dissociation is assumed, however, because the existence of these un-ionized gaseous species is poorly delineated and quantitative thermal data for them are lacking. The reactions which occur in the HI decomposition step are



At thermodynamic equilibrium, reaction 19 proceeds to only about 25% completion, so that a very large recycle of undecomposed HI (in the form of NH_4I) is unavoidable. Furthermore, the effect of temperature on equilibrium is weak. There is little to gain in trying to thermally shift the equilibrium until the temperature is dropped below the condensation point of elemental I_2 [normal boiling point of I_2 is 457.5°K (364°F)]. The gas-phase homogeneous HI decomposition is slow below about 700°K (590°F) and is virtually nil below 500°K (800°F), so that it is not practical to operate in a thermodynamically favorable temperature region unless some currently unknown catalyst for decomposition exists. About 98% of equilibrium decomposition is assumed.

Reaction 20 is a purely undesirable side reaction whose extent of occurrence in practical process operating circumstances is difficult to assess. This reaction is discussed further in Section 3.6.5.1.

The reaction which occurs in the ammonium iodide recovery step is assumed to be



Most of the ammonium iodide is recycled to the dissociation step, but a small portion of it is recycled to the ammonium iodide - ammonium sulfate synthesis step of the water splitting reaction subsystem to act as a catalyst. This catalyst NH_4I is accompanied by about 87% of the iodine produced by HI decomposition, so that the stream is actually most of the iodine recycled to the water splitting subsystem.

The ammonium iodide recovery step is also part of the hydrogen purification subsystem as well as the hydrogen production subsystem. The assignment of this step to the hydrogen production subsystem was made on the basis of its principal intended function, which is internal recycle of NH_4I to the dissociation step.

3.6.4.3. Hydrogen Purification Subsystem. The hydrogen purification subsystem consists of four principal processing steps:

1. Iodine recovery.
2. Ammonia recovery.
3. Ammonia resynthesis.
4. Hydrogen separation.

Figure 3-16 illustrates the material transfer relationships among these principal processing steps.

A gaseous mixture of NH_3 , H_2 , N_2 , and I_2 is received from the hydrogen production subsystem and is compressed first to $3 \times 10^6 \text{ N/m}^2$ (435 psi) and then to $2.5 \times 10^7 \text{ N/m}^2$ (3600 psi). The transformation which occurs in the iodine recovery step is



All but a trace of iodine vapor is recovered as solids by operating at $3 \times 10^6 \text{ N/m}^2$ (435 psi) and 325°K (125°F) effluent temperature. Minimizing the iodine content in the ammonia stream is not crucial in normal operation, but a severe leak of a water-filled heat exchanger into this part of the system would create a potential explosion hazard by the possible formation of nitrogen tri-iodide. This potentiality is discussed in more detail in Section 3.6.5.2.

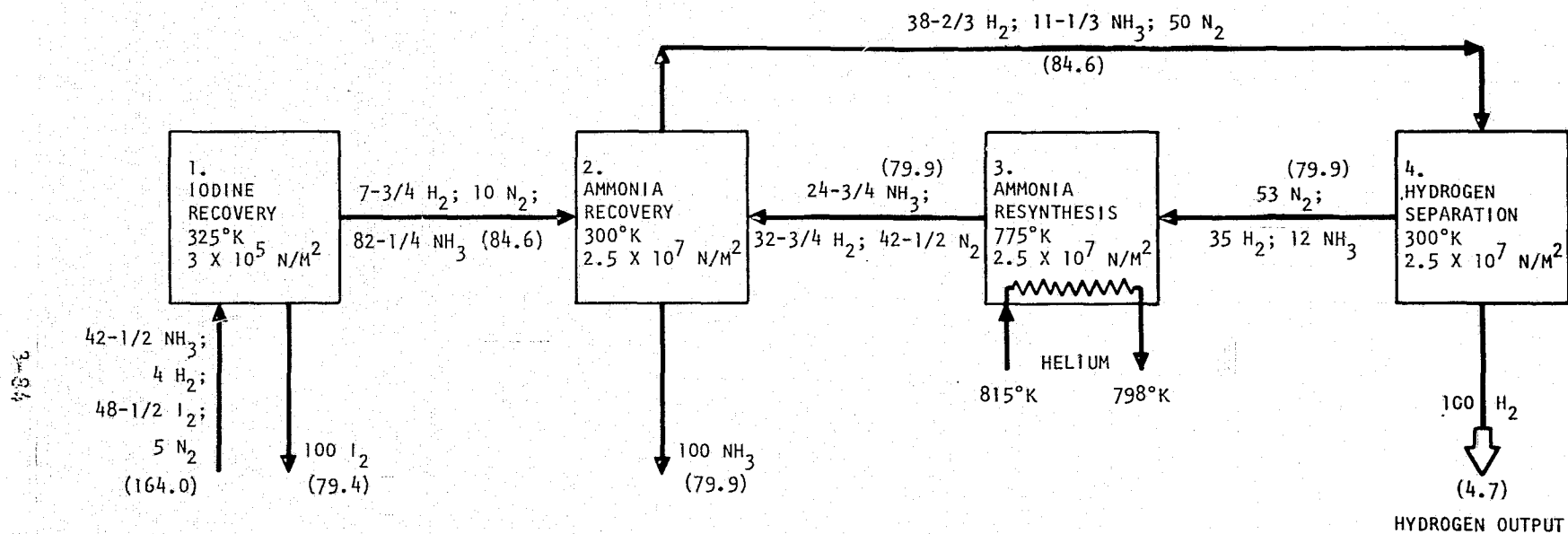
The transformation which occurs in the ammonia recovery step is



The fraction of total ammonia input liquefied in this step is about 90% for a condenser gas effluent temperature of 300°K (80°F). On the basis of only the ammonia which is returned to the condenser from the ammonia synthesis step, about 52% of the ammonia is liquefied.

The reaction which occurs in the ammonia synthesis step is the classical Haber process reaction



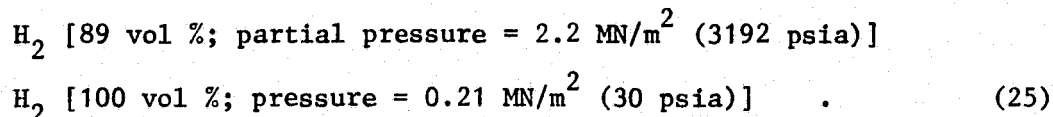


NOTES:
 NUMBERS ASSOCIATED WITH STREAM COMPONENTS ARE APPROXIMATE WT % COMPOSITIONS.
 NUMBERS IN PARENTHESES ARE APPROXIMATE TOTAL MASS FLOW RATES IN KG/SEC.

Fig. 3-16. Hydrogen purification subsystem

The amount of ammonia which has to be resynthesized is the amount which decomposes in the HI decomposer of the hydrogen production subsystem. Uncertainties in the amount of ammonia to be resynthesized are discussed further in Section 3.6.5.1.

The transformation which occurs in the hydrogen separation step is



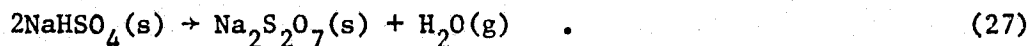
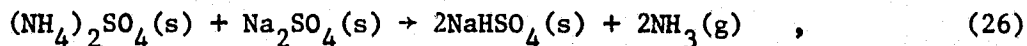
The method by which this transformation is hypothetically accomplished is diffusion through palladium. Unlike all the other unit operations hypothesized for this system, this separation method is unique for hydrogen. Furthermore, diffusional separation of gas constituents is not common or normally economical in large-scale chemical industrial practice. An almost unique example of large-scale diffusional separation is the production of U-235-enriched uranium, which is an ultra-high-cost product. However, the possibility of palladium diffusion for recovering pure H_2 has specifically been suggested, albeit informally, by Ispra workers, so the method is hypothetically incorporated into the Ispra Mark-10 process. Further discussion of hydrogen separation is presented in Section 3.6.7.1.

3.6.4.4. Oxygen Production Subsystem. The oxygen production subsystem consists of four principal processing steps:

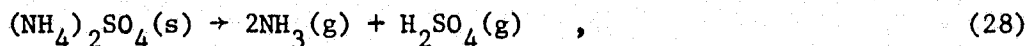
1. Sodium pyrosulfate synthesis.
2. Sodium pyrosulfate decomposition.
3. Sulfur trioxide decomposition.
4. Ammonium sulfate recovery.

Figure 3-17 illustrates the material transfer relationships among these principal processing steps.

Ammonium sulfate is received from the water splitting reaction subsystem and is mixed with sodium sulfate, which recycles internally in the oxygen production subsystem. The mixed salt solids are then heated to 675°K (775°F) in the pyrosulfate synthesis step to make sodium pyrosulfate by the following two reaction steps:

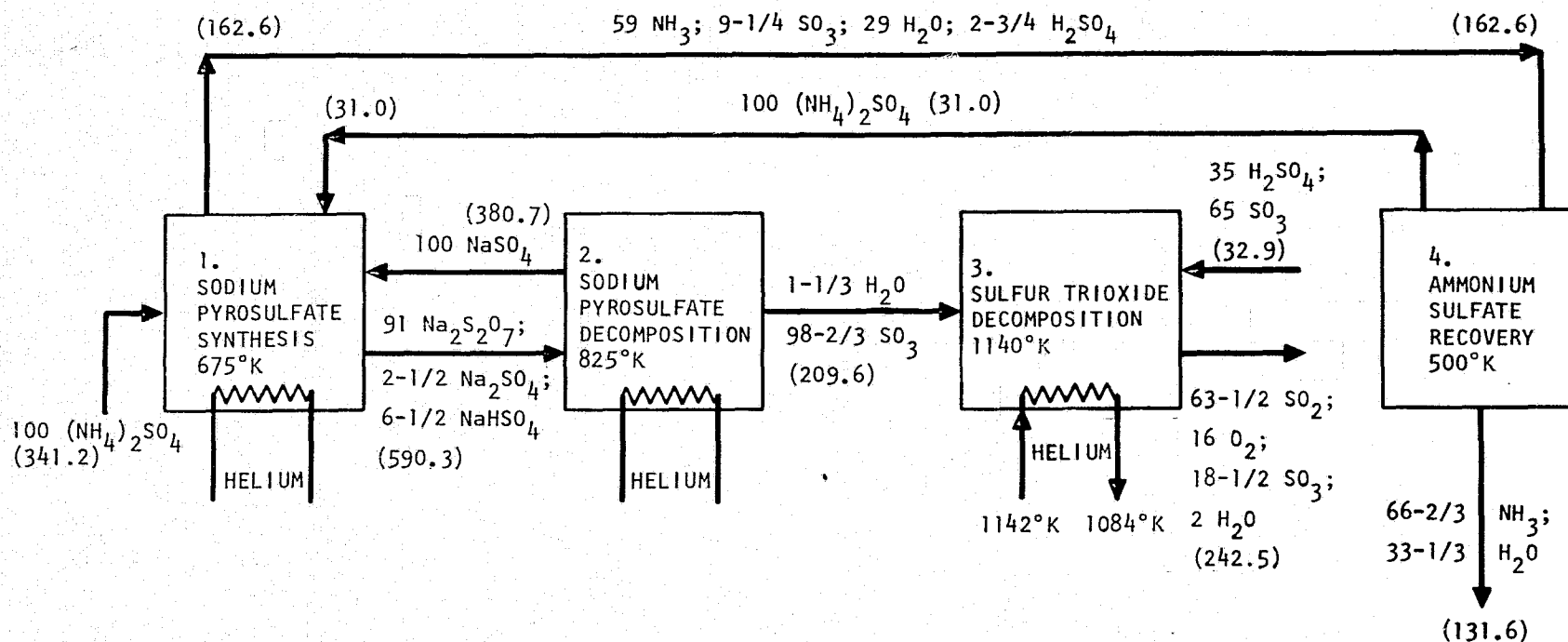


Reaction 26 appears to proceed with reasonable ease and completeness, but reaction 27 is difficult to push beyond about 90% to 95% completion under conditions which produce pyrosulfate. During the solid-state reactions 26 and 27, about 10% of the $(\text{NH}_4)_2\text{SO}_4$ typically commonly sublimes. It seems unlikely that molecular $(\text{NH}_4)_2\text{SO}_4$ would be the evolved gaseous species. It is assumed that the volatile species which sublime from heated $(\text{NH}_4)_2\text{SO}_4$ are NH_3 , H_2SO_4 , and the dissociation products (H_2O and SO_3) of H_2SO_4 . The reactions which express these assumptions are



The proportion of initial $\text{H}_2\text{SO}_4(\text{g})$ which is dissociated is about 80%.

Temperature control and intimacy of mixing of the salts are important for the operability of the pyrosulfate synthesis step. If the salt mix is overheated, a melt whose principle constituent is $\text{Na}_2\text{S}_2\text{O}_7$ forms, and the rate of reaction decreases drastically.



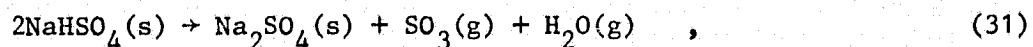
NOTES:
 SUBSYSTEM PRESSURE IS NOMINALLY $1.013 \times 10^5 \text{ N/M}^2$ THROUGHOUT.
 NUMBERS ASSOCIATED WITH STREAM COMPONENTS ARE APPROXIMATE WT % COMPOSITIONS.
 NUMBERS IN PARENTHESES ARE APPROXIMATE TOTAL MASS FLOW RATES IN KG/SEC.

Fig. 3-17. Oxygen production subsystem

The nature of the difficulty with the second reaction step cannot be described with great confidence, but it is probably a kinetic rather than a thermodynamic limitation. In general, solid-state reactions are typically subject to problems related to either diffusion of reactant molecules together or diffusion of reaction products out of the cake, or both.

The consequence of the returned water or equivalent of water in the sodium pyrosulfate is the generation of sulfuric acid when the effluent from the sulfur trioxide decomposition step is cooled. This sulfuric acid has only minor incidental advantages in a few process steps. Its main effects are to contribute an unproductive mass of material going through temperature cycling and to convert what otherwise would be useful high-temperature heat in the oxygen production system into waste heat of acid-base neutralization at 325°K (125°F) in the ammonium iodide - ammonium sulfate synthesis step.

The reactions which occur in the sodium pyrosulfate decomposition step are



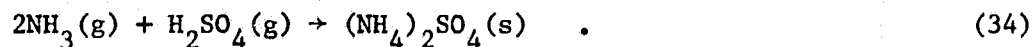
At 825°K (1025°F), the gas-phase recombination of SO_3 and H_2O proceeds to only a small extent, to about 3% of the water initially present.

The reactions which occur in the sulfur trioxide decomposition step are reactions 29 and 33:



Reaction 30 proceeds to about 80% completion (to near-thermodynamic equilibrium) at 1140°K (1592°F). Reaction 29 is virtually complete at 1140°K (1592°F), but since there is little $\text{H}_2\text{SO}_4(\text{g})$ present initially, the direct thermal consequences of this reaction in the SO_3 decomposition step are minor.

The reactions which occur in the ammonium sulfate recovery step are the reverse of the sublimation reactions which occurred in the pyrosulfate synthesis step; i.e., reactions 32 and 34:



Virtually all the gaseous sulfur-bearing species are condensed out as solid $(\text{NH}_4)_2\text{SO}_4$ at 500°K (440°F). Excess ammonia and water which accompany the dissociation products leave the subsystem, and the condensed $(\text{NH}_4)_2\text{SO}_4$ is recycled to the pyrosulfate synthesis step.

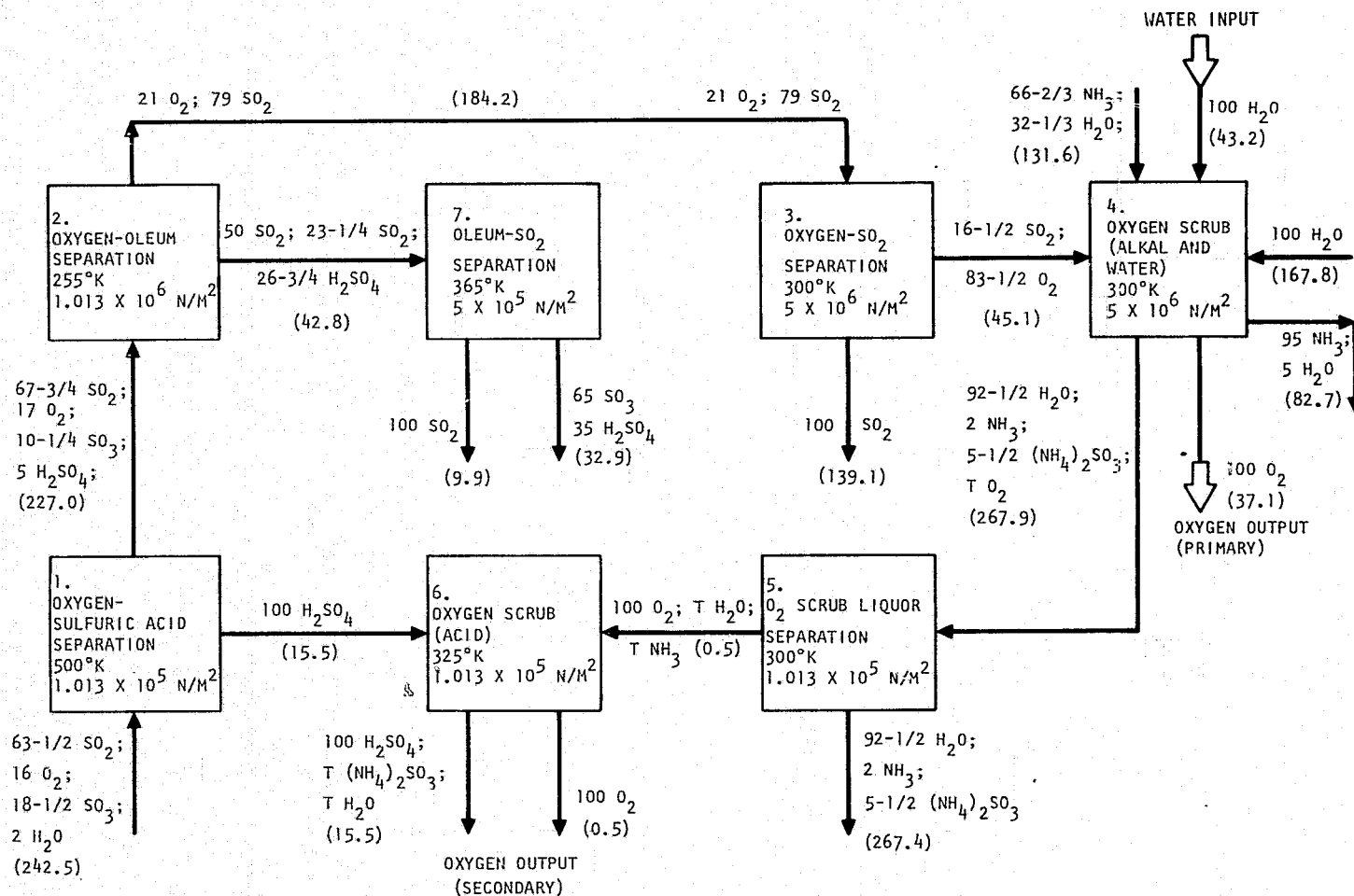
3.6.4.5. Oxygen Purification Subsystem. The oxygen production subsystem consists of seven principal processing steps:

1. Oxygen - sulfuric acid separation.
2. Oxygen-oleum separation.
3. Oxygen - sulfur dioxide separation.
4. Oxygen scrubbing (alkali and water).
5. Oxygen - scrub liquor separation.
6. Oxygen scrubbing (acid).
7. Oleum - sulfur dioxide separation.

Figure 3-18 illustrates the material transfer relationships among these processing steps.

Several alternatives to and variants of the particular processing steps chosen for incorporation into the oxygen purification subsystem can be conceived. For example, the flow diagram could be made to look simpler if the acidic scrub were omitted. Only about 1% of the total oxygen effluent appears in the stream which is acid scrubbed. The dissolved oxygen which separates from the alkaline scrub liquor when the pressure is reduced from $5 \times 10^6 \text{ N/m}^2$ (725 psi) to $1.013 \times 10^5 \text{ N/m}^2$ (147 psi) could simply be pumped back up into the alkali and water scrubber. But the apparent gain in simplicity by this change is more an artifact of the kinds of steps which are included or excluded from the highly schematic block diagrams rather than a real gain. In particular, pumps, compressors, power generators, and (with a single special exception) heat exchangers are not explicitly shown. The gas compressor for recycling ammonia-contaminated oxygen to the alkali and water scrubber would be more expensive to buy, maintain, and operate than a simple atmospheric-pressure scrub column.

A process step whose need is more questionable is the oleum- SO_2 separation step. It may be that the harm done by directly recycling a small amount of SO_2 (about 6% of the SO_2 produced in the SO_3 decomposition step) into the SO_3 decomposition step instead of into the $\text{NH}_4\text{I}-(\text{NH}_4)_2\text{SO}_4$ synthesis step is not worth the cost and operating expense of an SO_2 - SO_3 still and its attendant auxiliaries. This step is left in the flow diagram because the vapor pressure of SO_3 is so high at room temperature that semicryogenic conditions [about 255°K (0°F)] are needed to make a reasonably clean separation of SO_3 vapor from oxygen. Actual data for the phase relationships, solubilities, vapor pressures, and gas-liquid equilibrium composition are not available, and even the thermodynamic data for pure SO_3 and pure H_2SO_4 become sparse and of dubious character below room temperature. The problem is especially severe with SO_3 , which may exist in several solid allotropic forms at low temperatures. These forms may exist in mutual solid solution with each other. The composition of the



NOTES:

NUMBERS ASSOCIATED WITH STREAM COMPONENTS ARE APPROXIMATE WT % STREAM COMPOSITIONS.
NUMBERS IN PARENTHESES ARE APPROXIMATE TOTAL MASS FLOW RATES IN KG/SEC.

T = TRACE.

Fig. 3-18. Oxygen purification subsystem

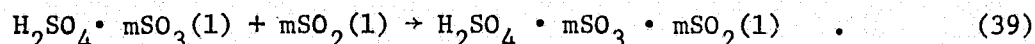
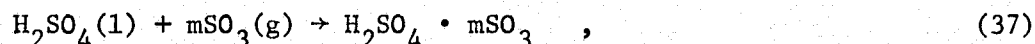
$\text{SO}_3\text{-H}_2\text{SO}_4\text{-SO}_2$ stream from the oxygen-oleum separation step is little better than a weak guess based on ideal solution theory. The SO_2 content may be off by a factor of two either way. If it is off by a factor of two low, an oleum- SO_2 separation step would probably be needed.

The transformations which occur in the oxygen - sulfuric acid separation step are



Under the assumed conditions, all the water is combined in the H_2SO_4 , in which relatively little SO_3 will be dissolved. Half of the condensed H_2SO_4 from this step is used to scrub ammonia from the secondary stream of oxygen product.

In the oxygen-oleum preparation step, the principal transformations are

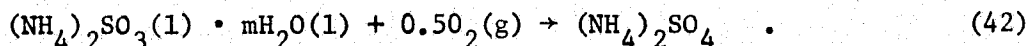
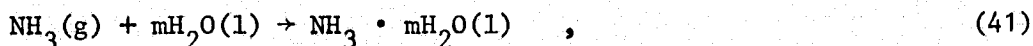
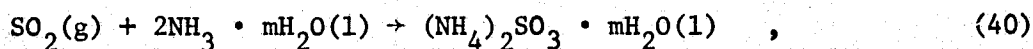


Some $\text{H}_2\text{SO}_4(\text{l})$ is introduced to help scrub the SO_3 from the vapor phase and to provide a liquid stream instead of a solid condensate of SO_3 from this step. Some SO_2 is also likely to be condensed from this step. The actual preparation of SO_2 condensed and in solution with the oleum is not supported by actual data for the properties of this ternary mixture. The composition given should be regarded as a reasonable assumption and not a fact based on data. One of the purposes of the oxygen-oleum separation

step is also to separate oleum from SO_2 . Some SO_2 in the oleum is desirable, because at 255°K (0°F) the oleum would be a solid unless some additional soluble liquid component were present. Exactly how much SO_2 would be needed to keep the condensate fluid is not known. Without the SO_2 , the freezing point of the $\text{H}_2\text{SO}_4 \cdot n\text{SO}_3$ would be in the vicinity of 275°K (-29°F). It is felt that having nearly 30 mole % of SO_2 [melting point = 197.7°K (-168°F)] would probably keep the mixture liquid at 255°K (0°F).

In the oxygen- SO_2 separation step, the principal transformation is reaction 38. About 95% of the SO_2 initially present in the gaseous O_2 - SO_2 mixture condenses under the temperature and pressure conditions [300°K (80°F), $5 \times 10^6 \text{ N/m}^2$ (725 psi)] selected.

In the oxygen scrub (alkali and water) step, the principal reactions are



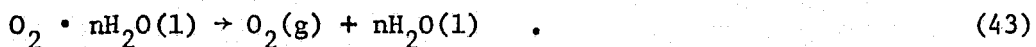
The function of this step is to remove the remaining 5 mole % SO_2 from the O_2 . A scrub with aqueous ammonia accomplishes this removal, but at the expense of putting more gaseous ammonia into the oxygen. The ammonia in the product oxygen is then scrubbed to some specified level with clean water.

The scrubbing operation might be slightly simplified by using only water to scrub the SO_2 from the oxygen, but a two-section scrub column, as assumed in this report, appears preferable. Ammonia is about five times as soluble in water as SO_2 on a weight basis (ammonia is about eighteen times as soluble on a mole basis), so the number of transfer units (NTU) needed

for ammonia is fewer. Furthermore, low levels of ammonia released to the environment are less noxious than equivalent amounts of low-level SO_2 .

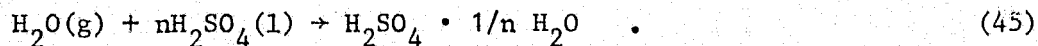
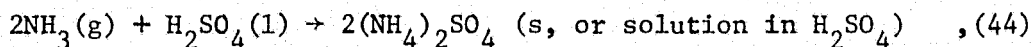
Oxygen - scrub liquor separation is introduced into the processing system to disengage dissolved oxygen from the scrub liquor. The fraction of oxygen appearing again as gas when the scrub liquor is depressurized is small, about 1%, but the absolute amount for the size of process plant which would couple with a 3000-MW(t) HTGR is 0.373 kg/sec during operation (almost 8000 tonnes/yr). It is hypothesized that this quantity of easily recoverable, near-pure oxygen would not be vented to waste.

The principal reaction which occurs in the oxygen - scrub liquor separation step is



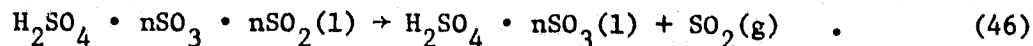
A small amount of ammonia accompanies the evolved oxygen. The function of the oxygen scrub (acid) step is to remove the traces of ammonia which accompany oxygen evolved in the oxygen - scrub liquor separation step. Water could be used, but since the recycle H_2SO_4 stream exists, it is used to make the ammonia removal easier and more complete. It would also be feasible to substitute a portion of the H_2SO_4 stream to replace the scrub water used in the oxygen scrub (alkali and water) step.

The principal reactions in the oxygen scrub (acid) step are



In addition to removing ammonia, H_2SO_4 acts as a strong desiccant in the gaseous stream.

The principal transformation which occurs in the oleum-SO₂ separation step is



In this step, which is envisioned as distillation, the H₂SO₄ is not expected to have any appreciable volatility. It should act as an extraction agent, which makes SO₂-SO₃ separation easier and more complete.

3.6.4.6. Connections Among the Subsystems. The overall process has considerable internal cross-connections and recycling. Any division into subsystems results in somewhat arbitrary assignment of certain processing steps to one or another of the subsystems. In addition, division into subsystems makes it difficult to adequately display the internally connected streams.

Figure 3-19 shows the steps of all the subsystems integrated into a single overall process system. The number which accompanies each stream is the rounded, approximate relative mass ratio of that stream with respect to the mass of hydrogen output.

3.6.4.7. Work and Heat Balance. With one special exception on Fig. 3-19, the preceding discussion and figures refer to only material transformations. The net energy loads for each of the major processing steps are shown in Table 3-13.

3.6.5. Side Reactions with Potential Process Significance

Some side reactions of the Ispra Mark-10 cycle are inherent in its chemistry, but many other possibilities are contingent on the completeness with which certain separations are made.

3.6.5.1. Inherent Side Reactions. A major uncertainty in the material balance is how much ammonia will crack in the HI decomposition step.

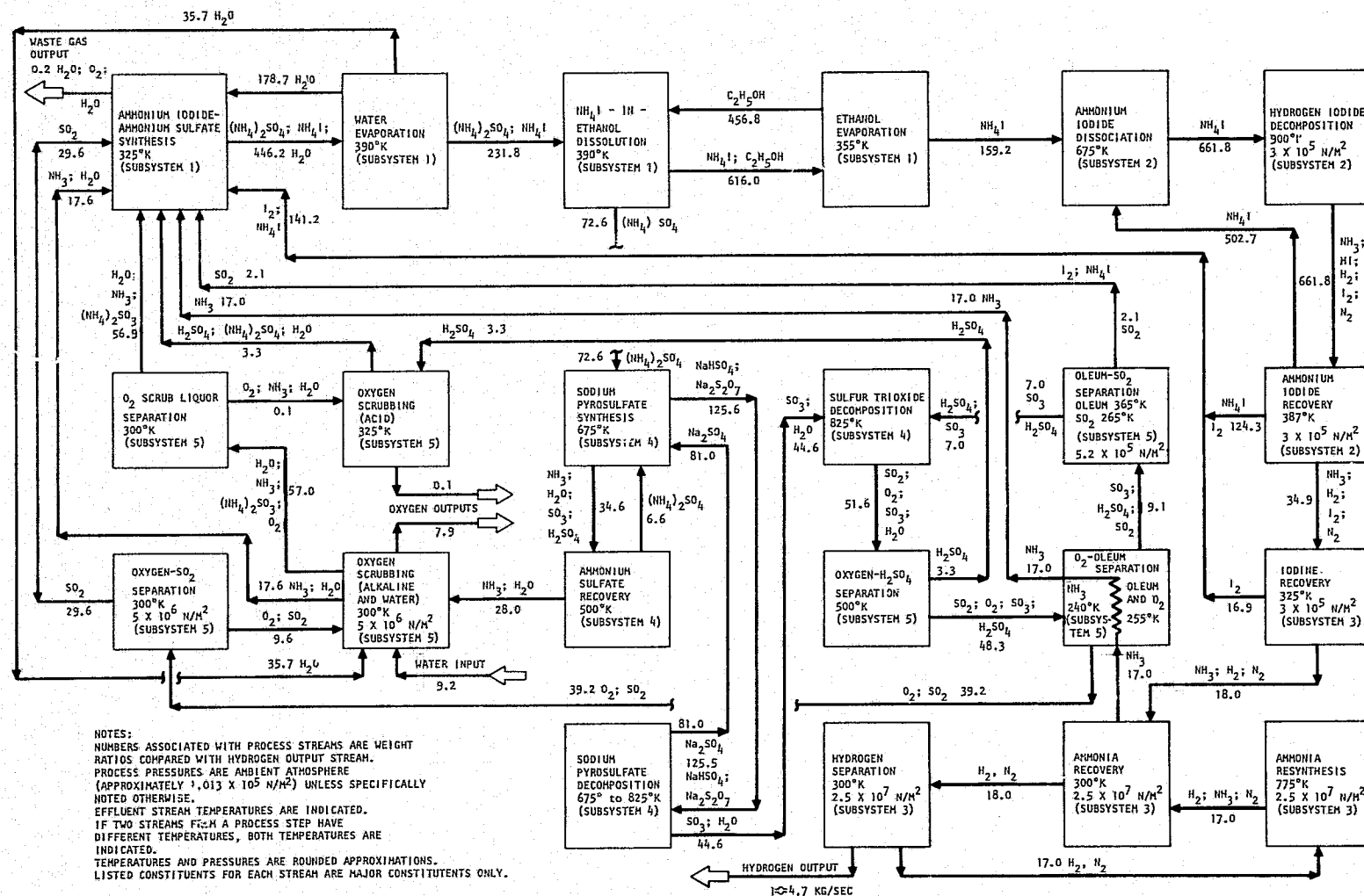


Fig. 3-19. Water splitting overall material balance

TABLE 3-13
NET ENERGY INPUTS AND OUTPUTS, ISBRA MARK-10 PROCESSING STEPS

Subsystem Number	Processing Steps	Direct Helium Heat In (MW)	Indirect Helium Heat In (MW)	Cooling Water Heat Out (MW)	Indirect Helium Heat Out (MW)
1	$\text{NH}_4\text{I}-(\text{NH}_4)_2\text{SO}_4$ synthesis	--	--	1192	--
1	Water removal	--	--	--	--
1	NH_4I in ethanol dissolution	--	--	88	--
1	Ethanol removal	--	--	690	--
2	NH_4I dissolution	--	1133	--	--
2	HI decomposition	62	--	--	--
2	NH_4I recovery	--	--	--	--
3	Iodine recovery	--	--	90	--
3	Ammonia synthesis	102	--	--	--
3	Ammonia recovery	--	--	102	--
3	Hydrogen separation	--	--	--	--
4	$\text{Na}_2\text{S}_2\text{O}_7$ synthesis	160	576	--	--
4	$(\text{NH}_4)_2\text{SO}_4$ recovery	--	--	111	--
4	$\text{Na}_2\text{S}_2\text{O}_7$ decomposition	131	--	--	--
4	SO_3 decomposition	361	--	--	--
5	$\text{O}_2\text{-H}_2\text{SO}_4$ separation	--	--	--	--
5	$\text{O}_2\text{-oleum}$ separation	--	--	38	--
5	$\text{SO}_2\text{-oleum}$ separation	--	--	--	--
5	$\text{O}_2\text{-SO}_2$ separation	--	--	25	--
5	$\text{O}_2\text{-scrub liquor}$ separation	--	--	--	--
5	O_2 scrub ($\text{NH}_4\text{OH-H}_2\text{O}$)	--	--	36	--
5	O_2 scrub (H_2SO_4)	--	--	--	--
-	Three-stage power generator(a)	--	--	--	1709
Subtotal		816	1709	2300	1709

TABLE 3-13 (continued)

	Direct Helium Heat In (MW)	Indirect Helium Heat In (MW)	Cooling Water Heat Out (MW)	Indirect Helium Heat Out (MW)
Subtotal (continued)	816	1709	2300	1709
Water decomposition energy and gas compression			700	
Power input from helium		<u>475</u>		
Total		3000 In	3000 Out	

(a) Net power generation of 475 MW is utilized entirely within the process. 700 MW leaves the process as energy of decomposition of water and as energy of compression of H_2 and O_2 product gases.

Homogeneous decomposition of pure ammonia is negligible even at temperatures of 1200° to 1300°K (1700° to 1880°F), but decomposition is catalyzed by the surfaces of metals and oxides commonly used to construct high-temperature/high-pressure process equipment. Iron, molybdenum, tungsten, and their oxides combined with silica and alumina catalyze the decomposition of ammonia.

Ispira experience with ammonia decomposition in the Mark-10 cycle has been inconsistent. The fraction of ammonia observed to decompose has ranged from low, perhaps about 2%, to virtually complete. The dissociation/decomposition products of NH_4I may have some homogeneous catalytic activity for ammonia decomposition.

In the material balance, 12% of the ammonia input (as ammonium ion) to the hydrogen production subsystem is assumed to decompose. This assumption is arbitrary, but a technically defensible value would require detailed, specific equipment design and more experimental work on the catalytic activity of the decomposition products from NH_4I with respect to ammonia decomposition.

The back-reaction of hydrogen and iodine, i.e.,



occurs to some extent during the ammonium iodide recovery step as HI is removed from the vapor by condensation of solid NH_4I . The amount of back reaction depends on how rapidly the decomposed gas traverses the temperature range of about 730° to 500°K (854° to 440°F). Condensation of NH_4I begins at about 730°K (845°F). Below 500°K (440°F), the kinetics of the recombination should be negligibly slow. However, at 730°K (854°F), the recombination rate is two-thirds the value of that at 900°K (1160°F).

Calculating a numerical value for the extent of back reaction involves detailed design of heat exchange equipment. No allowance for recombination

of H_2 and I_2 is included in the material balance. The temperature drop for heat transfer to cool the decomposed NH_4I in the NH_4I vaporizer is about $50^\circ K$ ($90^\circ F$), and a minimum ΔT of about $40^\circ K$ ($72^\circ F$) occurs in the equipment where the input plus recycle NH_4I first begin to vaporize.

3.6.5.2. Contingent Side Reactions. Most of the reactions discussed in this section depend on how completely some of the process separations can be made. The effects of any cooling water inleakage or corrosion are considered. With two exceptions, contingent side reactions are not specifically included in the material balances. These exceptions are

1. Incomplete removal of water from the pyrosulfate salt product from the pyrosulfate synthesis step.
2. Combinations of process constituents which could cause an explosion.

The first exception is a contingency only in that the amount of water remaining with the pyrosulfate salt is not certain. It is probable that this water in the pyrosulfate salt is present as $NaHSO_4$ rather than as free water which is physically trapped. This reaction is discussed in Section 3.6.4.5.

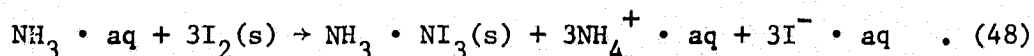
The second type of exception relates to the explosion possibilities of mixing hydrogen and oxygen gases and to the possibility of forming the explosive compound nitrogen tri-iodide (actually ammono-nitrogen tri-iodide, $NH_3 \cdot NI_3$).

Because of explosion hazards, hydrogen, oxygen, and iodine contents are quantitatively estimated at low flow rates and concentrations in the detailed material balance (Appendix A). In the absence of explosion hazards, constituent levels this low are called trace amounts and are without numerical specification.

Dissolved hydrogen and oxygen evolve back to the free gaseous state from process liquids which pass from higher pressure to lower pressure. In particular, hydrogen dissolved in liquid ammonia at $2.5 \times 10^7 \text{ N/m}^2$ (3600 psi) is released in the ammonium iodide - ammonium sulfate synthesis step and mixes with oxygen evolved by recycle streams from the oxygen purification system. The actual rates at which gaseous hydrogen and oxygen evolve in the ammonium iodide - ammonium sulfate synthesis step are uncertain because gas solubility data are found only for water. Gas solubilities in nonaqueous liquids are assumed to be about the same on a mole fraction basis as those in water.

The rates at which hydrogen and oxygen are carried into the ammonium iodide - ammonium sulfate synthesis step are about 0.024 kg/sec and 0.75 kg/sec, respectively, for a process sized to couple with a 3000-MW(t) HTGR. This hydrogen rate is only about 0.5% of the hydrogen production rate, but it takes only about 220 sec (less than 4 min) to accumulate a gaseous $\text{H}_2\text{-O}_2$ explosive mixture having an explosive energy content equivalent to 45.5 kg of sensitized fertilizer-grade ammonium nitrate blasting agent.

Nitrogen tri-iodide is formed when aqueous ammonia contacts elemental iodine:



As long as the $\text{NH}_3 \cdot \text{NI}_3$ is wet, it is not extremely sensitive to initiation of explosion, but when it is dry, it explodes if subjected to slight motion or friction.

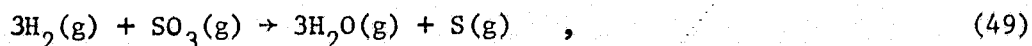
During normal process operation, iodine contacts aqueous ammonia only in the presence of SO_2 , and the iodine reacts with the SO_2 rather than the ammonia. However, abnormal but credible operating circumstances could lead to tri-iodide formation. An explosion in the water evaporation step could occur if the flow of SO_2 to the ammonium iodide - ammonium sulfate

synthesis step failed. Nitrogen tri-iodide might form if water from one of the heat exchangers downstream of the HI decomposition step leaked. Other credible explosion hazards are possible if more than one operating abnormality is hypothesized for a single incident.

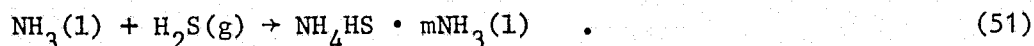
The separation steps which are most significant for generating side reactions in the process are separation of NH_4I from $(\text{NH}_4)_2\text{SO}_4$ and drying of the salt solids from water and ethanol.

Traces of NH_4I in $(\text{NH}_4)_2\text{SO}_4$ would be recycled to the ammonium iodide - ammonium sulfate step via pyrosulfate synthesis, ammonium sulfate recovery, oxygen scrubbing, and O_2 - scrub liquor separation steps. If the proportion of NH_4I in the $(\text{NH}_4)_2\text{SO}_4$ exceeds about 0.05 wt %, however, NH_4I would build up indefinitely in the recycle loop between the pyrosulfate synthesis and the ammonium sulfate recovery steps.

Traces of ammonium sulfate in the NH_4I solution in ethanol are reduced to either H_2S or elemental sulfur in the HI decomposer:



Any H_2S formed would go to the hydrogen purification system rather than the iodine recycle streams. The dissociation pressure of ammonium acid sulfide (NH_4HS) exceeds $1.013 \times 10^5 \text{ N/m}^2$ (147 psi) at temperatures above 306.5°K (92°F). Ammonium acid sulfide forms in the ammonia recovery step at 300°K (80°F):



The sulfide leaves the ammonia recovery step in the liquid ammonia stream and appears in the vapor from the water evaporation step. After the vapor is condensed, the H_2S can be removed by an air or nitrogen sparge if

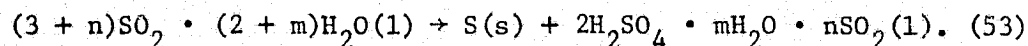
the condensate is neutral or acid. If the stream is alkaline, NH_4HS accumulates in the recycle water condensate loops to the ammonium iodide - ammonium sulfate step and the oxygen scrubbing step.

If elemental sulfur forms instead of H_2S , most of the sulfur accumulates in the NH_4I recycle loop of the hydrogen production subsystem. If some sulfur leaves the recycle loop in the liquid iodine stream, the sulfur does not accumulate indefinitely in the loop, but the equilibrium level of sulfur in the loop might be unacceptably high. The sulfur buildup may be controlled by adding a side stream of NH_4I solids going to aqueous dissolution and filtering. Filtering sulfur from a gaseous portion of the NH_4I recycle loop might be problematical because of the comparatively low melting point [386°K (235°F)] and boiling point [717.8°K (832°F)].

Sulfur reaching the iodine recycle streams is filtered out of the NH_4I -ethanol solution and goes with the $(\text{NH}_4)_2\text{SO}_4$ to the pyrosulfate synthesis step. The sulfur vaporizes, leaves with the ammonia-water vapor stream, passes through the ammonium sulfate condenser, and reappears as solids in the ammonia solution condenser. The sulfur solids then recycle back through the oxygen scrubbing step and appear again in the $(\text{NH}_4)_2\text{SO}_4$ salt. The sulfur builds up in this loop until its concentration in the $(\text{NH}_4)_2\text{SO}_4$ salt stream reaches 1.1 wt %, after which the sulfur inventory in the recycle loop remains constant and all additional sulfur passes as a melt with the sodium pyrosulfate stream to the pyrosulfate decomposition step. The sulfur is either oxidized to SO_2 or vaporized and then oxidized in the next step, SO_3 decomposition, at 1140°K (1592°F):



Elemental sulfur might also be produced if the heat exchanger in the oxygen- SO_2 separation step leaks:

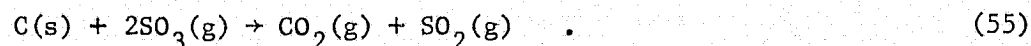


The free energy of this reaction is positive unless enough water is present for m to exceed about 1.5. At room temperature, the rate of disproportionation of the SO_2 is not extremely fast; sulfur solids usually require one or several hours rather than a few seconds or minutes to appear in a fixed-batch system.

Traces of ethanol in the $(\text{NH}_4)_2\text{SO}_4$ salt from NH_4I dissolution in ethanol vaporize in the pyrosulfate synthesis step and travel with the water vapor until they become trapped in the water condensate recycle loop of the ammonium iodide - ammonium sulfate synthesis step and the oxygen scrubbing step. The alcohol concentration builds up indefinitely, but the concentration of alcohol which can be tolerated in the water is probably high. Increasing alcohol in the water probably, at some point, precipitates $(\text{NH}_4)_2\text{SO}_4$ solids in the iodide - sulfate synthesis step and $(\text{NH}_4)_2\text{SO}_3$ in the oxygen scrubbing step. These precipitates would be a nuisance in equipment not designed to handle dilute slurries, but such equipment is not difficult to design nor expensive to make.

Traces of ethanol in the NH_4I salt from the ethanol evaporation step are reduced to carbon or hydrocarbons in the HI decomposition step. The carbon would then build up in the NH_4I recycle loop in the hydrogen production subsystem. Carbon might be removed by filtering the vaporized NH_4I stream [nominally $\text{NH}_3(\text{g}) + \text{HI}(\text{g})$] between the NH_4I evaporation step and the HI decomposition step.

If any carbon from ethanol leaves in the recycle iodine it appears as solids with the $(\text{NH}_4)_2\text{SO}_4$ salt and is ultimately converted to CO or CO_2 by hot SO_3 in the pyrosulfate decomposition step:



If the ethanol is converted to hydrocarbons in the HI decomposition step, the ultimate fate of the contained carbon becomes difficult to project. Light hydrocarbons might crack out pyrocarbon on the heat exchanger in the ammonia resynthesis step. Heavy hydrocarbons might condense with the liquid ammonia and collect in the heat exchanger for the O₂-oleum separation step.

The last kind of contingent side reaction considered is corrosion. The most corrosive materials in the Ispra Mark-10 process are sulfuric acid and iodine. Oleum, SO₃, HI, and SO₂ are also corrosive, and sodium acid sulfate and sodium pyrosulfate become corrosive in the presence of water. The pyrosulfate is slowly converted to the acid sulfate by moisture:



Other salt solids are considered noncorrosive or only mildly corrosive. Oxygenated water, as in the oxygen scrubbing step, is mildly corrosive.

Ammonia, nitrogen, and hydrogen are usually considered noncorrosive in a general sense, but they can sometimes attack certain types of construction materials under special circumstances. Hydriding, nitriding, and severe chilling can cause embrittlement and fracture of some metals. Hydriding is likely with platinum metals and group IV-A metals. Many steels and iron alloys become brittle when chilled.

The most severe corrosion problems are likely in the hot decomposition steps. Some reasonable candidate construction materials for these severe service applications exist, but direct experimental testing of candidate materials under projected service conditions could indicate a need for extensive material development.

High-silicon alloy steels, such as the Durirons, and silicon carbide, are candidate materials in sulfuric acid - SO₃ service. For iodine-iodide service, the high-silicon alloy steels are again candidates, along with

Hastelloy-B and, in the lower-temperature equipment, lead-lined steel. These suggestions are not exhaustive, especially for some of the lower-temperature, less severe service conditions. Noble-metal-lined and glass-lined equipment are candidates in some services, and the parts of the system under less severe service conditions with sulfuric acid and sulfates might be constructed of columbium or tantalum alloys.

The products of corrosion (metallic oxides and salts) are likely to accumulate wherever a solid stream is converted to a gas stream or in solid streams which are recycled. Places where such conditions occur are the ammonium iodide dissociation step, the boiling ammonia heat exchanger in the oxygen - oleum separation step, and the pyrosulfate decomposition step.

3.6.6. Materials Availability

The process materials for Ispra Mark-10 water splitting cycles can be supplied from water, ammonia, sulfuric acid, soda ash, ethanol, and iodine. Alternative source materials include sulfur dioxide, sulfur trioxide, ammonium sulfate, sodium sulfate, sodium acid sulfate, and caustic soda. Deionized water is the only material consumed in the process. Its cost varies widely depending on the quality of the raw water available for treatment. Deionized water costs in the range of \$0.25 to \$1.00/m³ (\$1.00 to \$4.00 per thousand gallons) might be typical.

U.S. production rates and prices for ammonia, sulfuric acid, soda ash, ethanol, and iodine in the mid to late 1960s are given below.

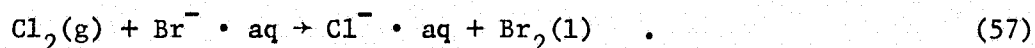
	<u>Production (kg/yr)</u>	<u>Price (dollars/kg)</u>
Ammonia	6×10^9	0.10
Sulfuric acid	2×10^{10}	0.017
Soda ash	5×10^9	0.031
Ethanol	2×10^9	0.19
Iodine	1×10^6 *	3.00

*World production of 9×10^6 kg/yr.

Iodine is the main problem when considering cost and availability. Under the conditions specified for HI decomposition, the iodine inventory of this step alone would be on the order of 4×10^6 kg for a process sized to couple with a 3000-MW(t) HTGR. The only obvious way to decrease this inventory is to increase temperature and pressure in the HI decomposition step. It seems especially appropriate to consider increased pressure, since the decomposition rate is proportional to the pressure. At 3×10^6 N/m² (435 psi) instead of 3×10^5 N/m² (44 psi), the iodine inventory needed in this step would be about 4×10^5 kg instead of 4×10^6 kg.

In the mid-1960s, Chile supplied about 50% of the world's iodine as a byproduct of saltpeter production from desert caliche deposits. Japan supplied 40%, mainly from the extraction of ashes from seaweed. The U.S. and the rest of the world supplied the remaining 10%. Most U.S. production comes from byproducts of oil well brines in California and Louisiana. These sources do not lend themselves to rapid expansion.

The possibility of direct extraction of iodine from seawater is attractive in some ways but problematical in others. Seawater contains 0.05 ppm of iodine. Bromine is economically recoverable from seawater at a concentration of 65 ppm for a price of \$0.070/kg (mid-1960s). The commercial recovery of bromine from seawater is based on the reaction



It is thermodynamically possible to drive the similar reaction for producing I₂(s) using either Cl₂ or Br₂.

The iodine supply problem is somewhat mitigated by its being for inventory rather than consumption. In normal operation, no iodine is consumed or lost. Iodine emission from the process would be strictly adventitious, abnormal, and hopefully rare.

3.6.7. Environmental Constraints and Plant Effluents

3.6.7.1. Normal Plant Effluents. In normal operation, the only effluents are pure hydrogen, nearly pure oxygen, and warmed cooling water. The hydrogen purity is assumed to be 99.99% based on a somewhat speculative suggestion made at Ispra: palladium permits the diffusion of hydrogen, but passes virtually nothing else. Commercially available palladium alloy thimble units for purifying small amounts of hydrogen have been on the market for years, but large-scale equipment based on this purification method has not been developed. The cost of palladium may preclude its use for large-scale purification of hydrogen despite its being a well known hydrogen purification method which has already been commercialized on a small scale.

The main oxygen output stream contains water vapor, a trace of ammonia, and possibly a trace of nitrogen. Placing a numerical value on the trace level of ammonia in the oxygen involves specifying the number of transfer units or the height of a transfer unit in the oxygen scrubbing column. The high solubility of ammonia in water^{*} permits its being scrubbed from the oxygen with reasonable ease. To scrub the oxygen to 50 ppm (the recommended occupation exposure threshold limit value in air) requires about six transfer units. Since the height of a transfer unit in typical NH_3 -water scrubbing columns is on the order of 1/3 to 2/3 m (1.1 to 2.2 ft), the total physical height of the scrubbing column which would reduce the ammonia in the oxygen to 50 ppm should be on the order of 4 m (13 ft).

The secondary oxygen stream is scrubbed by sulfuric acid and has an effluent temperature of 325°K (125°F). The vapor pressure-dissociation pressure of sulfuric acid at 325°K (125°F) is about 7.5 N/m^2 (1.09×10^{-3} psi), corresponding to about $4.7 \text{ mg H}_2\text{SO}_4/\text{m}^3$ in air. The recommended acceptance exposure threshold limit for H_2SO_4 is 1 mg/m^3 in air. This

^{*} 44 g NH_3 /100 g H_2O under an NH_3 pressure of $1.013 \times 10^5 \text{ N/m}^2$ (14.7 psi).

secondary oxygen stream is about 1% of the main oxygen stream. If it were combined with a small portion (about 1/4%) of the main oxygen stream, the H_2SO_4 - SO_3 vapors would react with the ammonia to give solid ammonium sulfate salt, which has little or no toxicity.

Although the foregoing impurity levels in the oxygen are low enough to meet occupational exposure standards, the total amount of impurities released in 1 yr is not small. From a water splitting process sized to couple with a 3000-MW(t) HTGR, the 9.31×10^8 kg of oxygen which would be produced (931,000 metric tons of O_2 /yr) would also carry out 4.6×10^6 kg/yr of water vapor, 45,000 kg/yr of NH_3 , and 3,800 kg/yr of $(\text{NH}_4)_2\text{SO}_4$. The cost of the material in the losses might total about \$5,000/yr at about 1965 prices.

The only normal material effluent which is wholly a waste stream is the waste gas vent stream (process vessel off-gas) from the ammonium iodide - ammonium sulfate synthesis step (see Fig. 3-14). Exclusive of any dilution gas addition to obviate explosion hazards, the normal material effluents in this stream are 1.9×10^7 kg of oxygen/yr, 5.9×10^5 kg of hydrogen/yr, and 2.4×10^6 kg of water vapor/yr. The higher heating value of the hydrogen is 8.3×10^{13} J/yr.

The normal waste having the most severe environmental impact is waste heat. A total of 7.6×10^4 TJ/yr of waste heat dissipated at a rate of 2270 MW while the process is running would be rejected to the environment. It is assumed that any desired amount of cooling water at 290°K (62°F) would be available.

Other assumptions about waste heat discharge are possible. It should be possible to increase the temperatures of the process streams from which the waste heat is removed, which are 300°K (80°F) and 325°K (125°F), by 30°K (54°F) or possibly even 40°K (73°F) before the ability to compensate in the other process conditions becomes a serious limitation.

3.6.7.2. Contingent Plant Effluents. Disastrous explosions or other major disruptions of a process plant are not discussed. The contingent effluents considered are mainly related to the contingent side reactions discussed in Section 3.6.5.

In normal operation and with clean separations, the waste gas vent stream contains only oxygen, hydrogen, and water (plus dilution gas, possibly). However, it takes a relatively small imbalance in the stream flows entering the ammonium iodide - ammonium sulfate synthesis step to put large amounts of ammonia or sulfur dioxide into this waste stream within a short time. If a severe imbalance in the direction of loss-of-SO₂-input occurs iodine vapor could appear in this waste stream. A very severe abnormality might lead to sublimation of NH₄I, or even more remotely to the generation of HI in the absence of ammonia. This latter severe abnormality would probably call for process shutdown. The other two less severe abnormalities might call for process shutdown in many, but not necessarily all, cases. In particular, carrying out iodide sulfate synthesis so that a small amount of SO₂ appears in the gas may be the safest way to operate.

The foregoing process upsets indicate the need for a versatile standby scrubber to handle conditions of mild upset in the iodide-sulfate synthesis step, especially during conditions not involving plant shutdown.

Some other gaseous products which might appear in the waste gas vent stream are CO, CO₂, ethanol vapor, and H₂S. The origins of these materials, which are contingent on particular conditions, are discussed in Section 3.6.5. Carbon monoxide is fairly toxic, but the concentrations involved in oxidizing traces of ethanol with the dried salts are negligible insofar as a practical CO poisoning hazard is concerned. Hydrogen sulfide is highly toxic as even very low concentrations and is even more toxic than hydrogen cyanide at the same concentration. When first smelled, hydrogen sulfide is detectable at levels well below the dangerously toxic level, but olfactory fatigue develops quickly and to a degree sufficient to allow a

lethal dose of H_2S to be inhaled without the victim being aware of the danger. The best way to handle an H_2S hazard is to avoid it by maintaining cleanliness when separating the $(NH_4)_2SO_4$ from the NH_4I . The cleanliness of this separation depends on keeping the ethanol nearly free of water.

The practical consequence of the requirement for near-anhydrous ethanol to ensure that no $(NH_4)_2SO_4$ dissolves is that a standby ethanol-water rectification column is needed. Mild upsets in the drying of the mixed salt from water evaporation, or even very long, continual normal operation, will lead to the accumulation of traces of water in the ethanol until levels sufficient to dissolve $(NH_4)_2SO_4$ are reached. A standby rectification system could also be used to keep any ethanol buildup in the water recycle loops within reasonable bounds.

3.6.8. Results and Discussion of Results

On the basis of the assumptions and limitations stated in Section 3.6.4, a water splitting process based on the Ispra Mark-10 reaction cycle was developed. This process converts 23% of the input heat* from a 3000-MW(t) HTGR into potential enthalpy of combustion of gaseous hydrogen at $2.1 \times 10^5 \text{ N/m}^2$ (30 psi) and 300°K (80°F). The hydrogen output design rate is 4.7 kg/sec while the process is operating. Annual production would be $1.19 \times 10^8 \text{ kg}$ of H_2 /yr at 80% onstream factor, limited by the HTGR being at 100% full power for 80% of the year.

The process efficiency of 23% is somewhat low, partly because of a desire not to depart very far from process conditions resembling those reported in the experiments at Ispra for developing the Mark-10 cycle and partly because of several vapor recycle streams whose energy of condensation is large.

*Higher heating value.

The most serious and inherent difficulty is the large recycle of NH_4I . Not more than about 25% of the HI present in the HI decomposer decomposes on a single run because of thermodynamic equilibrium limitations. Thus, it is necessary to recycle four times as much NH_4I as the net input of NH_4I from the synthesis step. Furthermore, the enthalpy of dissociation is about 1.16×10^6 J/kg of NH_4I , so that a very large amount of energy must be recovered and passed through a heat transfer temperature drop, which degrades the energy's availability for work or chemical free energy as H_2 .

The equipment engineering problems associated with the recovery of heat in the product gas from the HI decomposition step will be difficult. The vapor compressor between the NH_4I dissociation step and the HI decomposition step must operate on a fairly corrosive gas (50 vol % HI) under severe temperature conditions. The development of practical engineering equipment for countercurrent heat recovery from condensing NH_4I solids at an overall log mean ΔT in the vicinity of 50°K (90°F) will also be difficult.

A somewhat less serious but similar problem occurs during the recycle of Na_2SO_4 solids between the pyrosulfate decomposition step and the sodium pyrosulfate production step. Although the magnitude of the heat load in the pyrosulfate production step is significantly lower than it is in the HI dissociation step, accurate definition of its magnitude is hampered by a lack of accurate, relevant thermodynamic data for the main transformations involved. The heat loads in the pyrosulfate synthesis steps are considered to be rough estimates rather than hard data.

The heat loads in the water evaporation and the ethanol evaporation steps are kept within reasonable bounds by vapor compression and recovery of the heat of condensation at temperatures high enough for heat transfer back into the evaporators. Vapor compression heat recuperation is, however, capital and power intensive. The amount of water and ethanol specified by Ispra for dissolving the salts appears to be convenient rather

than optimized amounts. The potential for considerable heat, power, and capital savings may exist for the evaporation steps if the salt concentrations in the solvents are maximized, consistent with process operability.

The Ispra Mark-10 process has six principal solids handling steps. A very rough rule of thumb is that handling a given mass of solids costs about 1.5 times as much as moving the same mass of fluid.

The degree to which the nominal overall heat conversion efficiency might be raised by altering various assumptions about the process is theoretically limited only by the Carnot efficiency, with which the reactor heat could be converted to free energy (work). The thermodynamically ideal efficiency for the heat assumed in Section 3.6.4 to be supplied by the reactor and for heat rejection from process streams at 300° to 325°K (80° to 126°F) is nearly 0.7 [i.e., 2100 MW free energy from 3000 MW(t) total heat]. If all this work were used in a thermodynamically ideal water splitting process to make H_2 at 300°K (80°F) and $2.1 \times 10^5 \text{ N/m}^2$ (30 psia) and 0.50_2 at 300°K (80°F) and 3 MN/m^2 (435 psia), the ratio of required free energy (work) input to total energy input (enthalpy) would be 0.853*. The 900 MW of heat left over from work generation would then supply 360 MW of heat to the process to make up the difference between the total energy input required (enthalpy) and the work input required (free energy). The remaining 540 MW of heat must be rejected by the process strictly for thermodynamic reasons. Under the assumptions stated in Section 3.6.4., the actual heat rejection for the process is 2300 MW, of which 1760 MW is due to practical rather than theoretical thermodynamic limitations.

There are some methods for reducing other waste heat loss due to practical problems. One way is to minimize the amount of solvent H_2O and solvent ethanol which must be recycled. Another way is to reexamine and

*For $P_{H_2} = P_{O_2} = 1.013 \times 10^5 \text{ N/m}^2$ (147 psi) at 298°K (77°F), this ratio is 0.830.

more carefully reoptimize some of the assumed pressure-temperature conditions. Still other ways are to split and redirect some of the heating-cooling fluid recycle streams among various points of the process or to totally decompose the NH_3 of the NH_4I instead of recycling NH_4I solids. It should be possible to increase overall process efficiency to a value exceeding 30% (i.e., to decrease the 1760 MW of heat loss due to practical considerations to less than 1560 MW). However, to push the overall process efficiency to 40% less than 1260 MW heat loss due to practical conditions) would probably entail radical and unforeseeable improvements in the HI decomposition and NH_4I recycle steps.

An example of the kind of technological breakthrough that might make 40% efficiency a reasonable and practical goal is the development of a catalyst to permit operating the HI decomposition step at temperatures low enough for liquid I_2 to be withdrawn from the reaction mixture. The critical temperature of iodine is 819°K (1014°F) versus the Ispra specification of about 900°K (524°F) as the decomposition temperature. If I_2 can be removed from the mixture during decomposition, it should be possible to shift the reaction equilibrium to greater than 25% decomposition of the HI and thus reduce the enormous NH_4I recycle burden.

3.6.9. Data Sources

The sources for data on material properties and thermochemistry used in this report are Refs. 3-15 through 3-27. Cost and production data for materials are found in Refs. 3-22 and 3-28.

The data on thermodynamic properties and material properties are generally adequate for S, SO_2 , SO_3 , H_2SO_4 , H_2 , N_2 , I_2 , H_2S , H_2O , HI, and Na_2SO_4 as pure substances. Exceptions to this general adequacy are data on SO_3 and H_2SO_4 below room temperature. A large part of the problem with chilled SO_3 is that the solid can exist in several allotropic forms. These forms may occur simultaneously in mutual solid solution with each other and

with slowly shifting near-metastability. The data for NH_4I are not wholly concordant, and the exact identity of all gaseous species upon vaporization is doubtful.

Thermochemical data for $(\text{NH}_4)_2\text{SO}_4$ are restricted to near room temperature and are sparse compared with what is available for NH_4I . Quantitative data for $\text{Na}_2\text{S}_2\text{O}_7$ are even more sparse than data for $(\text{NH}_4)_2\text{SO}_4$. The little that is available in Ref. 3-5 is fragmentary and not subject to cross-checking for reliability. The heat effects calculated for process steps involving pyrosulfate are best regarded as very approximate, order-of-magnitude estimates.

Except for aqueous solutions of a single solute near room temperature, thermochemical property data for solutions are not available. Various methods are used to estimate properties of solutions in nonaqueous media and with more than one solute. Raoult's Law, Henry's Law, the Perfect Gas Law(s), including Dalton's Law and Amagat's Law, were assumed as needed. The linear additivity of partial molal properties upon mixing binary solutions was assumed. Vapor pressure data were sometimes extrapolated assuming a constant enthalpy of evaporation. Where heat capacity data were needed but not available, e.g., for ammonium sulfate at elevated temperature and for sodium pyrosulfate, quantitative estimates were generated by drawing analogies to similar compounds for which data are available; however, some inconsistencies may have developed in the crude analogies and proportionalities used. In some cases, estimates of properties are based on methods given in Ref. 3-29.

The most severe data inadequacies occur with solids rather than liquids or gases. Most of the correlation methods in Ref. 3-29 are developed for organic vapors and liquids, in which neither large solvation energies nor strong electrostatic effects dominate, as they often do in inorganic substances.

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4. HYDROGEN MARKET SURVEY

4.1. CURRENT CONSUMPTION

Domestic consumption of hydrogen for all uses was 99.1 billion m³ (3500 billion ft³) in 1972 and has been increasing at a rate of approximately 12%/yr in recent years. End uses for hydrogen and their growth rates in 1972 are shown in Table 4-1. As can be seen from this table, the petroleum refinery processes of hydrotreating, hydrocracking, and hydrodesulfurization have been consuming a proportionately larger share of the hydrogen produced in recent years. These refinery operations have become important because of the increasing attention being given to reduction of air pollution due to sulfur compounds and diminishing supplies of good quality fuels.

There is little published information on the quantities of hydrogen obtained from various sources in the U.S., but it is believed that petroleum refining operations, particularly catalytic cracking and catalytic reforming, supplied 45% of the hydrogen requirements in 1972. Reforming of natural gas supplied another 45% of the hydrogen, and the remaining 10% came from other sources such as byproducts of coal and coke manufacture, fuel oil refining, water electrolysis, and byproduct gases from ethylene manufacture or other chemical plants.

Most of the hydrogen produced in the U.S. is captive hydrogen; i.e., it is consumed either where it is produced or in nearby chemical plants as either a raw material or a fuel. A small percentage of the hydrogen produced is known as merchant hydrogen and is produced for sale by industrial gas companies.

TABLE 4-1
HYDROGEN MARKET CONSUMPTION AND GROWTH DURING 1972

End Use	Consumption			Recent Growth Rate (%/yr)
	% of Total	10 ⁹ M ³	10 ⁹ Ft ³	
Ammonia	37	36.5	1290	4
Hydrotreating and hydrodesulfurization	23	22.9	810	15
Hydrocracking	19	19.0	670	25
Chemical	9	8.8	310	9
Methanol	7	7.1	250	9
Miscellaneous	5	4.8	170	7
Total	100	99.1	3500	

Most merchant hydrogen shipments are accomplished in cylinders; however, a small portion is shipped by pipeline. Air Products currently operates an 80.5-km (50-mile) hydrogen pipeline that serves ten customers in the Houston area. They are considering a large partial oxidation unit that would process refinery hydrocarbon residues and return hydrogen gas. Liquid Carbonic is also supplying hydrogen by pipeline to consumers in the Geismar, Louisiana, area. The longest hydrogen pipeline in the world is currently in the Ruhr area of Germany. It is about 204 km (127 miles) long and has been in operation since the late 1930s.

Hydrogen plants associated with ammonia plants and petroleum refineries tend to be concentrated in a few regions. For example, there were about 100 ammonia plants in 31 states and Puerto Rico in 1968, with a total capacity of over 16.4 tonnes (17 million short tons)/yr. These plants are primarily located in the Midwest, the Gulf Coast area, western Texas, and California. Table 4-2 lists these plants and shows that the major concentration of ammonia production is in southern Louisiana, where 3683 tonnes (3950 thousand short tons) of the annual capacity are located. Large petroleum refinery centers are located on the Gulf Coast near Philadelphia and New York, on the west coast of California and Washington, and inland in the Chicago and St. Louis areas. U.S. petroleum refineries are listed by state in Table 4-3. Figure 4-1 shows that the U.S. refinery capacity is concentrated in ten regions, each with a refinery capacity of more than 5,565,000 m³/day (350,000 Bbl/day).

TABLE 4-2
 DIRECTORY OF CHEMICAL PRODUCERS, U.S.

<u>Ammonia</u>		<u>Annual Capacity</u> <u>(Thousands of Tons)</u>
Air Products and Chems., Inc.	New Orleans, La.	210
	Pensacola, Fla.	75
Allied Chem. Corp.		
Specialty Chems. Div.	Hopewell, Va.	340
	South Point, Ohio	340
Union Texas Petroleum Div.		
Agricultural Dept.	Geismar, La.	340
	Omaha, Neb.	200
American Cyanamid Co.		
Agricultural Div.	New Orleans, La.	340
Apache Powder Co.	Benson, Ariz.	15
Baker Indust. Corp.		
Agricultural Products Corp., subsid.	Conda, Idaho	100
Borden Inc.		
Borden Chem. Div.		
Petrochems.	Geismar, La.	285
CF Indust., Inc.		
Chattanooga Nitrogen Complex	Tyner, Tenn.	165
Donaldsonville Nitrogen Complex	Donaldsonville, La.	720
Fremont Nitrogen Complex	Fremont, Neb.	50
North Carolina Nitrogen Complex	Tunis, N. C.	210
Terre Haute Nitrogen Complex	Terre Haute, Ind.	135
Coastal States Gas Corp.		
Colorado Interstate Corp., subsid.		
Wycon Chem. Co., subsid.	Cheyenne, Wyo.	145
Colombia Nitrogen Corp.	Augusta, Ga.	130
Cominco American Inc.		
Camex, Inc., subsid.	Borger, Tex.	400
Commercial Solvents Corp.	Sterlington, La.	340
Cooperative Farm Chems. Association	Lawrence, Kans.	340
Diamond Shamrock Corp.		
Diamond Shamrock Oil and Gas Co.	Dumas, Tex.	160

TABLE 4-2 (Continued)

<u>Ammonia</u>		<u>Annual Capacity (Thousands of Tons)</u>
Dow Chem. U.S.	Freeport, Tex.	115
E. I. du Pont de Nemours & Co., Inc.		
Biochems. Dept.	Belle, W. Va.	340
Elastomer Chems. Dept.	Beaumont, Tex.	340
Plastics Dept.	Victoria, Tex.	100
El Paso Natural Gas Co.		
El Paso Products Co., subsid.	Odessa, Tex.	115
Farmland Indust., Inc.		
	Dodge City, Kans.	210
	Enid, Okla.	400
	Fort Dodge, Iowa	190
	Hastings, Neb.	140
Felmont Oil Corp.	Olean, N. Y.	85
First Mississippi Corp.		
FIRSTMISS INC., subsid.	Fort Madison, Iowa	340
FMC Corp.		
Chem. Group		
Indust. Chem. Div.	South Charleston, W. Va.	25
Gardinier, Inc.		
U.S. Phosphoric Products	Tampa, Fla.	130
Goodpasture, Inc.	Dimmitt, Tex.	85
W. R. Grace & Co.		
Agricultural Chems. Group	Big Spring, Tex.	125
	Memphis, Tenn.	330
Green Valley Chem. Corp.	Creston, Iowa	35
Gulf + Western Indust., Inc.		
The New Jersey Zinc Co., subsid.	Palmerton, Pa.	40
Hercules Inc.		
Synthetics Dept.	Hercules, Calif.	70
	Louisiana, Mo.	70
Kaiser Aluminum & Chem. Corp.		
Kaiser Agrigultural Chems. Div.	Savannah, Ga.	150
Lone Star Gas Co.		
Nipak, Inc., subsid.	Kerens, Tex.	125
	Pryor, Okla.	105

TABLE 4-2 (Continued)

<u>Ammonia</u>		<u>Annual Capacity</u> <u>(Thousands of Tons)</u>
Mississippi Chem. Corp.	Pascagoula, Miss.	175
	Yazoo City, Miss.	395
Mobil Oil Corp.		
Mobil Chem. Co., div		
Petrochems. Div.	Beaumont, Tex.	260
Monsanto Co.		
Monsanto Agricultural	Luling, La.	420
Products Co.		
N-Ren Corp.		
Cherokee Nitrogen Div.	Pryor, Okla.	55
High Plains Div.	Plainview, Tex.	25
Occidental Petroleum Corp.		
Hooker Chem. Corp., subsid.		
Hooker Chems. and Plastics		
Corp., subsid.		
Electrochemical and	Tacoma, Wash.	25
Chems. Div.		
Occidental Chem. Co.,		
subsid.		
Southern Region	Plainview, Tex.	55
Western Div.	Hanford, Calif.	25
	Lathrop, Calif.	95
Olin Corp.		
Agricultural Chems. Div.	Lake Charles, La.	485
Pennwait Corp.		
Chem. Div.	Portland, Ore.	10
Phillips Pacific Chem. Co.	Kennewick, Wash.	155
Phillips Petroleum Co.		
Fertilizer Div.	Beatrice, Neb.	210
Plastics Div.	Pasadena, Tex.	230
PPG Indust., Inc.		
Chem. Div.		
Indust. Chem. Div.	Natrium, W. Va.	50
Reichhold Chems., Inc.	St. Helens, Ore.	90
Rohm and Haas Co.		
Rohm and Haas Texas Inc.,	Deer Park, Tex.	50
subsid.		
St. Paul Ammonia Products,	East Dubuque, Ill.	210
Inc.		

TABLE 4-2 (Continued)

<u>Ammonia</u>		<u>Annual Capacity (Thousands of Tons)</u>
J. R. Simplot Co. Minerals and Chem. Div.	Pocatello, Idaho	110
Skelly Oil Co. Hawkeye Chem. Co., subsid.	Clinton, Iowa	140
Standard Oil Co. of California	El Segundo, Calif.	20
	Richmond, Calif.	135
Chevron Chem. Co., subsid. Ortho Div.	Fort Madison, Iowa	105
Standard Oil Co. (Kentucky), subsid.	Pascagoula, Miss.	500
Standard Oil Co. (Indiana) Amoco Oil Co., subsid.	Texas City, Tex.	700
The Standard Oil Co. (Ohio) Vistron Corp., subsid. Chems. Dept.	Lima, Ohio	500
Tenneco Inc. Tenneco Chems., Inc. Organics and Polymers Div.	Houston, Tex.	210
Tennessee Valley Authority	Muscle Shoals, Ala.	65
Terra Chems. Internat'l, Inc.	Port Neal, Iowa	210
Tipperary Corp.	Lovington, N. M.	35
Triad Chem.	Donaldsonville, La.	370
Tyler Corp. Atlas Powder Co.,	Joplin, Mo.	135
Union Oil Co. of California Collier Carbon and Chem. Corp., subsid.	Brea, Calif.	270
	Kenai, Alas.	520
United States Steel Corp.	Cherokee, Ala.	175
USS Agri-Chemicals, div.	Clairton, Pa.	325
	Geneva, Utah	70
Valley Nitrogen Producers, Inc.	El Centro, Calif.	210
	Helm, Calif.	175
Arizona Agrochemical Co., subsid.	Chandler, Ariz.	35
Vulcan Materials Co. Chems. Div.	Wichita, Kans.	25
The Williams Companies Agrico Chem. Co., subsid.	Blytheville, Ark.	340
	Donaldsonville, La.	340
Total		17,445

TABLE 4-3
SURVEY OF OPERATING REFINERIES IN THE U.S. (STATE CAPACITIES AS OF JANUARY 1, 1975)

State	No. Plants	Crude Capacity ^(a)		Charge Capacity, (bbl/stream day)								Production Capacity, (bbl/stream day)				Coke (tons/day)
		(bbl/calendar/day)	(bbl/stream day)	Vacuum Distillation	Thermal Operations	Cat Cracking		Cat Reforming	Cat Hydro-cracking	Cat Hydro-refining	Cat Hydro-treating	Alkylat-ion	Aromatics/ Isomerization	Lubes	Asphalt	
						Fresh Feed	Recycle									
Alabama	3	34,375	37,947	11,000	--	--	--	1,200	--	--	5,000	--	--	--	7,700	--
Alaska	4	66,050	69,520	--	--	--	--	--	--	--	--	--	--	--	300	--
Arizona	1	4,000	4,211	2,500	--	--	--	--	--	--	--	--	--	--	--	--
Arkansas	4	60,715	62,536	24,900	--	15,000	3,000	5,750	--	--	13,100	4,500	--	2,550	7,350	--
California	36	1,900,640	1,985,140	933,850	465,883	485,611	124,200	471,339	319,622	139,244	662,822	90,388	14,490	20,800	94,560	15,168
Colorado	3	60,000	62,500	19,500	29,600	21,600	700	12,000	--	--	16,300	--	--	--	3,300	57
Delaware	1	140,000	150,000	90,700	44,000	62,000	15,000	42,000	17,000	--	110,000	8,000	--	--	--	1,500
Florida	1	5,700	6,000	3,400	--	--	--	--	--	--	--	--	--	--	3,000	--
Georgia	2	18,000	19,789	--	--	--	--	--	--	--	--	--	--	--	9,000	--
Hawaii	2	85,000	89,473	15,000	--	14,100	8,900	--	--	--	1,400	4,130	1,350	--	1,300	--
Illinois	11	1,168,150	1,224,658	417,499	147,300	406,277	68,300	315,377	66,500	108,000	474,243	100,322	10,100	5,600	38,100	4,053
Indiana	8	563,275	595,973	264,200	32,500	187,100	35,000	123,500	--	29,500	184,000	28,800	6,500	7,900	36,400	860
Kansas	11	447,180	466,129	126,300	45,200	162,500	42,750	94,200	3,000	3,000	107,000	38,800	3,400	4,000	18,150	1,425
Kentucky	3	164,000	168,900	68,000	4,000	64,500	1,000	30,100	--	--	42,000	--	18,500	--	23,500	--
Louisiana	19	1,729,575	1,802,149	497,342	137,833	618,278	61,900	377,583	78,500	114,500	412,851	134,389	26,800	24,850	39,880	5,730
Maryland	2	26,500	28,211	13,800	--	--	--	--	--	--	--	--	--	--	19,800	--
Michigan	6	149,082	151,379	42,000	--	39,500	9,500	27,350	--	17,700	27,700	4,700	2,000	--	11,650	--
Minnesota	3	199,300	204,000	87,000	23,000	68,900	7,700	30,100	--	20,000	72,100	11,500	--	--	42,000	1,300
Mississippi	5	289,500	304,737	156,000	6,700	70,500	6,350	70,700	62,000	--	41,450	9,200	6,000	--	--	320
Missouri	1	107,000	107,000	40,000	11,000	41,000	12,000	13,000	--	--	52,000	4,500	--	--	6,500	--
Montana	8	157,206	164,227	50,250	11,850	45,300	25,300	36,050	5,020	--	103,400	10,200	5,100	--	23,425	250
Nebraska	1	5,000	5,500	2,400	--	2,400	500	1,100	--	--	--	--	--	--	--	--
New Jersey	4	539,000	562,764	286,633	38,144	224,444	40,000	93,944	--	50,000	229,945	17,133	--	6,400	73,000	975
New Mexico	7	103,061	111,135	12,400	1,250	12,400	5,160	11,570	--	--	28,550	2,925	--	--	700	--
New York	2	111,385	114,500	43,000	--	41,500	6,000	24,000	--	20,000	41,500	2,800	3,000	--	18,000	--
North Dakota	3	58,658	60,263	--	1,100	23,000	11,000	10,200	--	--	12,600	2,900	--	--	--	--
Ohio	7	589,770	614,500	207,500	27,700	204,460	44,240	165,000	82,000	45,000	159,000	35,300	--	21,000	19,600	1,270
Oklahoma	12	499,815	515,145	152,813	51,900	182,000	41,950	123,300	4,500	--	150,900	44,450	15,550	10,200	23,200	1,620
Oregon	1	14,000	14,740	15,000	--	--	--	--	--	--	--	--	--	--	8,600	--
Pennsylvania	11	757,020	796,415	328,378	26,750	205,000	18,300	221,708	52,700	173,000	278,250	38,100	10,200	29,928	36,500	--
Rhode Island	1	7,500	10,000	--	--	--	--	--	--	--	--	--	--	--	6,600	--
Tennessee	1	43,900	44,800	13,000	--	13,500	--	--	--	--	--	4,000	--	--	8,000	--
Texas	45	3,929,430	4,090,965	1,295,041	316,588	1,227,166	258,705	1,001,242	153,167	314,500	1,409,143	230,651	196,176	94,322	64,900	6,227
Utah	6	143,000	117,068	34,800	8,000	52,400	18,000	23,300	1,000	5,500	18,500	10,450	4,550	--	4,700	300
Virginia	1	53,500	54,000	28,000	14,000	27,000	5,000	9,000	--	--	24,000	--	--	--	--	710
Washington	7	364,000	379,876	133,016	36,000	91,500	27,100	79,888	35,000	20,500	155,667	21,333	2,900	--	7,000	1,500
West Virginia	3	14,750	20,500	8,675	--	--	--	6,160	--	4,440	7,510	--	--	6,770	--	--
Wisconsin	1	45,000	46,400	15,500	--	9,700	1,000	10,000	--	5,800	10,000	1,200	--	--	12,000	--
Wyoming	12	186,670	194,660	66,726	4,444	58,778	15,300	31,294	--	16,644	55,894	7,860	1,500	1,510	14,817	139
Total	259	14,845,407	15,463,850	5,497,143	1,484,742	4,677,414	933,855	3,461,955	880,009	1,087,328	4,906,825	868,511	328,916	216,930	683,532	43,404

(a) State totals include figures converted to calendar-day or stream-day basis.

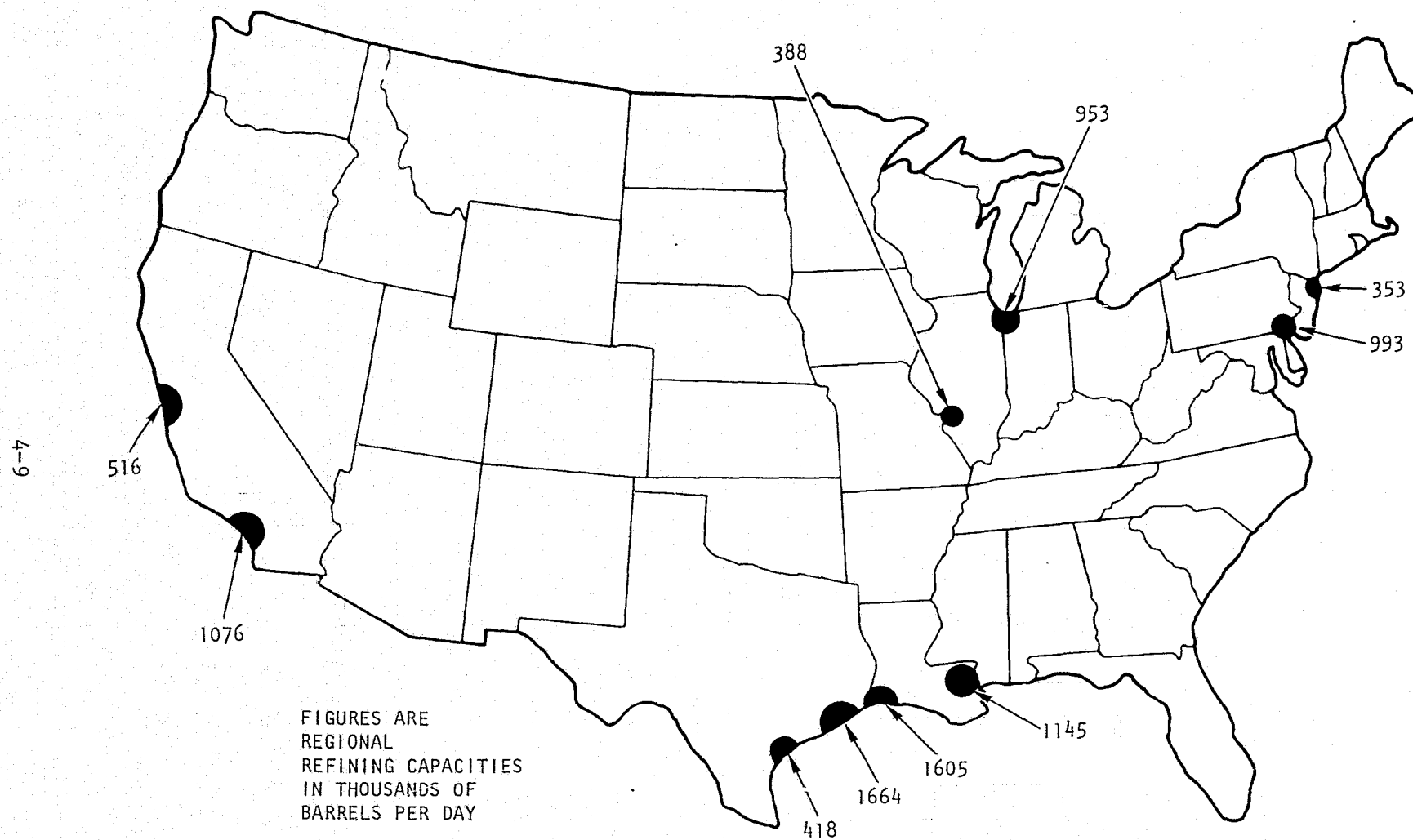


Fig. 4-1. Refining regions with a total capacity exceeding 5,565,000 m³/day

4.2. FUTURE DEMAND FOR HYDROGEN

The analysis of the future demand for hydrogen is approached on an end-use basis. The end uses covered are ammonia, petroleum refinery operations, chemicals, methanol, and other uses.

4.2.1. Ammonia

Forecasted annual increases in ammonia production for the next decade range from 2% to 7%. This spread in growth rates is caused by the uncertainties in forecasting agricultural demand, which consumes almost 80% of the ammonia produced in the U.S. The lower predicted growth rate is based on the assumption that the U.S. fertilizer market is now approaching saturation and further increases will be related to increases in population. However, fertilizer demand could get a big boost if some of the world's big granary countries, e.g., the U.S.S.R., the U.S., France, or Argentina, were to have poor harvests. The resulting low grain reserves could push fertilizer demand higher because more land would be brought into production.

On the supply side, ammonia producers are plagued by getting enough natural gas feedstock, which is the base feedstock for most of the country's ammonia plants. It has been estimated that 308,000 tonnes (340,000 short tons) of ammonia production will be lost owing to natural gas curtailments this spring. This problem is being approached by switching ammonia plants from natural gas to other fuels for heat sources. It is likely that other hydrocarbons will be evaluated as a feedstock for hydrogen manufacture by either steam reforming or partial oxidation. Consideration is also being given to the manufacture of ammonia from coal, starting with a coal gasification process to make product gas.

Another uncertainty in ammonia supply is the position of the Middle Eastern countries with regard to the large volumes of natural gas currently

being flared there. For example, Saudi Arabia has announced plans for increasing ammonia production by 1.8 million tonnes (2 million short tons)/yr by 1980. However, the lack of trained people and political unrest in this area could cause delays in these plans. If and when this ammonia capacity is realized, it more than likely will be used to satisfy the European market and will not be a factor in the U.S. market. Since ammonia in the U.S. is currently in tight supply, the U.S. has no export market, and anything happening in the European market will have no effect on the U.S. market.

4.2.2. Petroleum Refinery Operations

The petroleum refinery operations of hydrotreating, hydrodesulfurization, and hydrocracking, consumed $41.9 \times 10^9 \text{ m}^3$ (1.480 billion ft^3) of hydrogen in 1972, which is 42% of the hydrogen consumed in the U.S. Hydrogen processing has become an increasingly important part of petroleum refinery operations in recent years. More of the crude processed must be hydrogenated either to protect catalysts in subsequent processes, such as in catalytic reformers and hydrocrackers, or to produce the low-sulfur products required to meet environmental restrictions. Although the maximum amount of sulfur allowed in fuel oils for existing burners will probably level off after the provisions of the Clean Air Act of 1973 become fully effective, the average sulfur content of fuel oils will continue to decrease as more new furnaces subject to federal "new source" standards are built and old furnaces are retired.

With the decrease in the sulfur content of fuels, high-sulfur Middle Eastern crude is becoming a growing fraction of the crude processed in U.S. refineries at the expense of domestic crude. This trend is expected to continue at least until 1980. After 1980, the sulfur content of the average crude charged to the U.S. should decrease as 0.8%-sulfur Alaskan crude replaces some of the 1.8%-sulfur Middle Eastern crude. The average sulfur content of U.S. crude would decrease more rapidly if 0.02%-sulfur

synthetic crude oil from Canadian or Venezuelan tar ponds replaces some of the high-sulfur Middle Eastern crude oil in the U.S.

The typical U.S. refinery is in balance with respect to hydrogen; i.e., the hydrogen produced is equal to the hydrogen consumed. It is only when a refinery adds a hydrocracking unit or a chemical unit (such as ammonia) that a refinery is put out of balance and a hydrogen plant must be built.

The major sources of hydrogen in a refinery are fluid catalytic cracking (FCC) units used to upgrade heavy distillates to gasoline and catalytic reforming units used to convert naphthas into high-octane gasoline. The purity of hydrogen from an FCC unit is about 30% to 40%, and that from a catalytic reformer is about 80%.

It is expected that FCC units will retain their role as the prominent process for converting heavy distillates to gasoline because of recent improvements in technology. It is also expected that hydrocracking, used to convert distillates to jet fuels, naphthas, and motor gasoline blending components, will not change much beyond its current capacity level because economics favor FCC units.

The major growth in the use of hydrogen in refineries will be in desulfurization processes to meet sulfur specifications. There is a great variation in the hydrogen requirements among the petroleum fractions that are hydrogenated, primarily because of the sulfur that must be removed. Hydrogen consumptions range from 0.035 to 0.59 m³ H₂/m³ oil (30 to 50 scf/bbl) for naphtha through 0.118 to 0.236 m³/m³ (100 to 200 scf/bbl) for middle distillate to as high as 1.26 m³/m³ (700 scf/bbl) for residual fuel oil. As a rule of thumb, the typical U.S. refinery without a hydrocracking unit consumes about 0.236 m³/m³ (200 scf/bbl) of crude charged. This could increase to 0.354 m³/m³ (300 scf/bbl) as the amount of sulfur in the crude and/or the amount of product desulfurization increases.

In order to estimate the amount of hydrogen that will be required for petroleum refinery operations by the year 1980, an estimate of refinery capacity must be made. In 1972, U.S. refineries processed 186 million m^3 (11.7 million bbl)/day of crude. It has been forecast that this will increase to about 270 million m^3 (17 million bbl)/day by the year 2000. Assuming that the hydrogen requirements for a refinery are 0.54 m^3/m^3 (300 scf/bbl) of charge, this would mean a hydrogen requirement of 0.0526 trillion m^3 (1.86 trillion ft^3) of hydrogen.

4.2.3. Methanol

Methanol manufacture consumes only 7% of the hydrogen produced in the U.S., and this market is currently growing at a rate of 9%/yr. About 60% of the methanol produced in the U.S. is used for the manufacture of formaldehyde. It does not appear that the growth rate for methanol will change much in the next decade. Considerable publicity has been given to using methanol as a fuel, but this does not appear to be feasible, primarily because of the economics of manufacturing. Like the manufacture of ammonia, the manufacture of methanol is based on natural gas as a feedstock, and the current projected prices of natural gas make methanol prices uneconomic when compared with prices of other fuels. However, there reportedly are environmental advantages for using methanol as a fuel, and these advantages may become a significant factor in the future use of methanol.

4.2.4. Other Uses

Although the area of other uses is the one in which hydrogen has the most potential growth, it is also the area with the most uncertainties. Other uses include using hydrogen for steel manufacturing, coal gasification, and production of synthetic fuels from coal and oil shale. Hydrogen has also been proposed for use in transportation, but this application is not considered in this report.

4.2.5. Steelmaking

The use of natural gas to generate reducing gas for direct reduction of iron ore in steel manufacture is now a commercial process. Presently, there are 1.2 tonnes/yr (1.3 million short tons/yr) of steel manufactured by direct reduction in the U.S. It has been estimated by the American Iron and Steel Institute (AISI) that the use of direct reduction for steel manufacture could grow to the point that 184 billion m^3/yr (6500 billion ft^3)/yr of reducing gas (CO , H_2) would be required from 1985 to 2000. Since hydrogen and carbon monoxide are equivalent in their ability to act as reducing agents, this demand for reducing gas could be met by hydrogen alone.

4.2.6. Synthetic Fuels

The production of synthetic fuels from coal and tar sands would result in significant requirements for hydrogen. The U.S. Bureau of Mines has estimated that $0.37 \times 10^{12} \text{ m}^3$ (13 trillion ft^3) of pipeline quality gas could be produced from coal in the year 2000 to supplement natural gas supplies. Since one process to gasify coal requires 44.0 m^3 (1560 ft^3) of hydrogen to produce one million Btu of gas [approximately 28.3 m^3 (1000 ft^3)], it would take $574 \times 10^9 \text{ m}^3$ (20,300 billion ft^3) of hydrogen to produce this amount of synthetic gas. The U.S. Bureau of Mines has also estimated that $0.0283 \times 10^{12} \text{ m}^3$ (1 trillion ft^3) of high-Btu gas may be produced from oil shale by the year 2000. The conversion of oil shale by one process requires 1200 m^3 (1200 ft^3) of hydrogen per 1000 m^3 (1000 ft^3) of gas produced. On this basis, $33.97 \times 10^3 \text{ m}^3$ (1200 billion ft^3) of hydrogen would be required for this application. Thus, the potential requirements for hydrogen in the manufacture of synthetic gas from coal and oil shale will be $0.608 \times 10^{12} \text{ m}^3$ (21.5 trillion ft^3) in the year 2000.

Whether or not this market for synthetic fuels develops will depend, of course, on the economics of the process and the nature of U.S. national energy policy. It does not appear likely, however, that these plants will

represent a market for merchant hydrogen. Rather, it is more likely that these plants will consume captive hydrogen produced by partial oxidation of a residual stream available within the plant.

4.2.7. Chemicals

The manufacture of numerous chemicals and petrochemicals consumed $8.77 \times 10^9 \text{ m}^3$ (310 billion ft^3) of hydrogen in 1972. By far, the largest use of hydrogen in this application is for the manufacture of cyclohexane from benzene. Cyclohexane is primarily used in the manufacture of nylons 6 and 6/6. It has been estimated that consumption of cyclohexane will increase at 6%/yr up to the year 2000.

4.3. SUMMARY AND CONCLUSIONS

Hydrogen markets can be summarized as shown in Table 4-4. These data show that the major growth area for hydrogen will be in the production of synthetic fuels from coal and oil shale. Whether or not this market develops in the manner shown will depend on several uncertainties. These include the nature of the proposed national energy policy and whether new, large gas and petroleum reserves are discovered.

At the present time, it appears that these markets could be satisfied by mostly captive hydrogen plants rather than by merchant supplies. This is not to discount the possibility of the development of a large merchant hydrogen market. However, this will depend on economic and institutional factors. Merchant hydrogen producers must be able to sell large amounts of hydrogen at lower prices than can be produced by captive plants. Likewise, the producers must also be able to overcome the institutional problems expected to exist when large amounts of hydrogen are moved from producing to consuming plants; these problems have been solved in three geographic areas.

TABLE 4-4
FUTURE HYDROGEN MARKET

Application	Production [10^{12} m ³ /yr (10^{12} ft ³ /yr)]				
	1972		2000	%/yr	
Ammonia	0.0365	(1.29)	0.0637 - 0.243	(2.25 - 8.58)	2-7
Refining	0.0420	(1.48)	0.0526	(1.86)	1
Chemicals	0.0088	(0.31)	0.0447	(1.58)	6
Methanol	0.0071	(0.25)	0.0470	(1.66)	7
Miscellaneous	0.0048	(0.17)	0.0320	(1.13)	7
Steelmaking			0.184	(6.5)	--
Synthetic fuels			0.608	(21.5)	--
Total	0.0992	(3.50)	1.032 - 1.211	(36.5 - 42.8)	8.8 - 9.4

N76-15582

5. PROCESS DESCRIPTION

5.1. INTRODUCTION

The production of hydrogen from coal by hydrogasification and subsequent steam hydrocarbon reforming has been selected for further study. The basic process is described in Section 3.2 (process 2). Process heat and power requirements are supplied by an HTGR.

The study described in this section examines the process as a function of a maximum reforming (process) temperature of 922°K (1200°F) to 1367°K (2000°F). The major parameters for the five cases selected are summarized in Table 5-1. A fixed thermal capacity of 3000 MW for the HTGR heat source was used as a basis. The resulting hydrogen production rates and the corresponding hydrogen purity are also given in Table 5-1. The HTGR heat source for each of the five cases was the subject of a previous study (Ref. 5-1).

5.2. PROCESS DESCRIPTION

Process flow sheets were developed for the five cases covering process temperatures of 922°K (1200°F) to 1367°K (2000°F). The coal gasification process steps (Sections 100 to 500) were appropriately scaled for the five cases considered while maintaining identical process steps and process conditions. A description of the coal gasification plant is given in Section 3.2 and is not discussed further in this section. The hydrogen plant, which consists of hydrogen production and power generation and hydrogen purification and compression (Sections 600 to 700), required significant changes over the range of reforming temperatures considered. Case II, for a process temperature of 1033°K (1400°F), is described in

TABLE 5-1
SUMMARY OF MAJOR PARAMETERS
(3000-MW HTGR)

	Case I	Case II	Case III	Case IV	Case V
Process temperature (reformer) (°K) (°F)	922 1200	1033 1400	1144 1600	1256 1800	1367 2000
Reforming pressure (MN/m ²) (psia)	2.06 300	2.06 300	2.76 400	3.45 500	4.13 600
Steam/carbon mol ratio	4.5:1	4.5:	3:1	1.5:1	1.5:1
Hot helium temperature (°K) (°F)	1033 1400	1144 1600	1256 1800	1367 2000	1477 2200
Cold helium temperature (core inlet) (°K) (°F)	644 700	683 770	773 932	839 1050	922 1200
Hydrogen production rate (100% capacity) (kg/sec) (MMscfd) (a)	15.32 530	22.47 763	25.99 882	27.95 949	32.16 1092
Hydrogen purity	95.4 ^(b)	87.9	90.8	91.0	93.8
Export power (electric) (MW)	138	57	31	18	7

(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

(b) Case I utilizes a cryogenic unit for the upgrading of hydrogen purity.

detail as the base case. For other cases, only the required changes in the process are discussed.

In all cases, process heat and power requirements are supplied by a 3000-MW HTGR. The HTGR nuclear heat source for each case was taken from previous studies (Ref. 5-1). In all cases, the helium loop conditions and designs given in Ref. 5-1 were maintained. Small differences in reformer feed composition required minor changes in reformer process gas flow rate and reformer process gas inlet temperature to match HTGR helium loop conditions and HTGR designs.

5.2.1. Case II (Base Case), Process Temperature = 1033°K (1400°F)

A summary of plant performance is given in Table 5-2. Additional details on power requirements and cooling water requirements are given in Tables 5-3 and 5-4. Overall stream flows are given in Figs. 5-1 and 5-2 and Tables 5-5 and 5-6.

5.2.1.1. Hydrogen Production and Power Generation (Section 600). The following description refers to the process flow sheet shown in Fig. 5-3 and the mass balance given in Table 5-7.

Helium, which serves as the primary reactor coolant, carries heat from the HTGR nuclear core to the reformers and steam generators. Helium entering the nuclear core at 683°K (770°F) is heated to 1144°K (1600°F). The hot helium supplies heat for the reforming reaction, leaving the reformers at 933°K (1219°F). The helium stream then transfers additional heat to the steam generators, leaving at 669°K (744°F). Helium is compressed and returned to the core. The distribution of HTGR thermal power is as follows:

TABLE 5-2
OVERALL PERFORMANCE SUMMARY, CASE II, PROCESS TEMPERATURE = 1033°K (1400°F)

	Quantity at 100% Capacity	Gross Heat Value		% of Total Input
		(MW)	(10 ⁹ Btu/hr)	
Input				
Coal feed	186.71 kg/sec (17,780 tpd)	5410	18.47	64.3
Makeup water	217.14 kg/sec (3443 gpm)			
HTGR	3000 MW	3000	10.24	35.7
Total in		8410	28.71	100.0
Product				
Hydrogen	12.692 kg-mole/sec (868 MMscfd) 87.9 mol % ^(a)	4405	15.04	52.5
Aromatics	18.93 kg/sec (10,300 bpd)	808	2.76	9.5
Power	57 MW	57	0.19	0.7
Subtotal		5270	17.99	62.7
Waste heat				
Char	34.11 kg/sec (3250 tpd)	636	2.17	7.5
To cooling water		2311	7.89	27.5
Other		193	0.66	2.3
Subtotal		3140	10.72	37.3
Total out		8410	28.71	100.0

(a) At 273°K (32°F), 0.101 MN/m² (14.7 psia).

TABLE 5-3
POWER REQUIREMENTS, CASE II
PROCESS TEMPERATURE = 1033°K (1400°F)

	Power (MW)
Section 600	
Power produced	
Circulator turbine	95
High-pressure turbine	209
Low-pressure turbine	129
Subtotal	433
Power consumed	
Helium circulator	(95)
Pumps	(18)
Subtotal	(113)
Net power produced, Section 600	320
Section 700	
Power produced	--
Power consumed	
Hydrogen compression	(227)
Pumps	(12)
Other	--
Subtotal	(239)
Net power consumed, Section 700	(239)
Net power consumed (Sections 100 to 500)	(24)
Total net power produced (Sections 100 to 700)	57

TABLE 5-4
COOLING WATER REQUIREMENTS, CASE II,
PROCESS TEMPERATURE = 1033°K (1400°F)

Item No.	Description	Heat Duty	
		MW	10 ⁹ Btu/hr
Sections 600, 700			
T-604	Power steam condenser	281	0.959
T-704	Condenser	156	0.533
T-706	Condenser	195	0.666
T-707	CO ₂ stripper condenser	819	2.796
T-708	Condenser	59	0.235
R-701	Compressor intercoolers	141	0.482
Total Sections 600 to 700		1661 MW	(5671 MMBtu/hr)
Total Sections 100 to 500		650 MW	(2219 MMBtu/hr)
Total heat duty		2311 MW	(7890 MMBtu/hr)

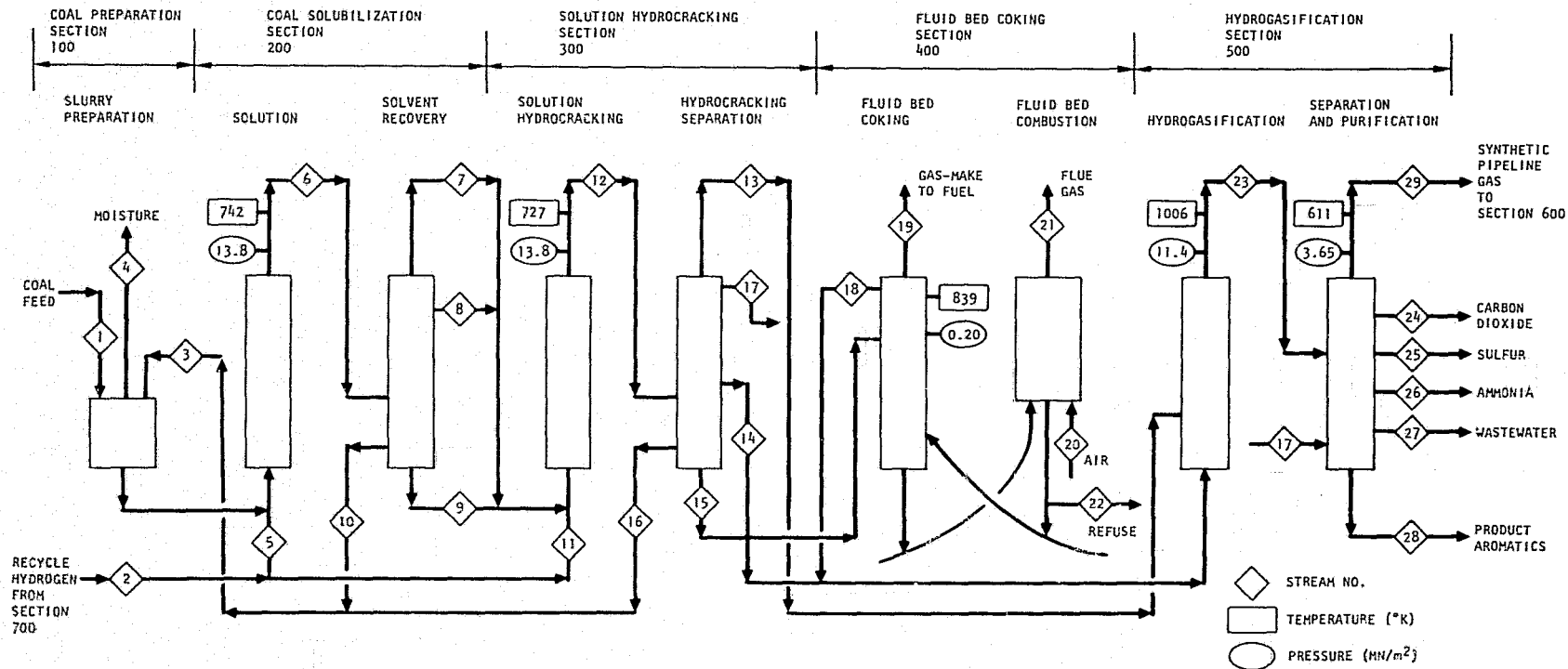


Fig. 5-1. Flow diagram, Sections 100-500, Case II

TABLE 5-5

HYDROGASIFICATION OF COAL LIQUIDS OVERALL PLANT FLOWS, CASE II, PROCESS TEMPERATURE = 1033°K (1400°F)

Stream No.	1	2	3	4	5	6	7	8	9	10
Component Description	Coal Feed	Recycle Hydrogen	Recycle Solvent	Moisture	Hydrogen to Coal Solution	Coal Solution Effluent	Hydrocarbon Gases	Distillate	Coal Liquids	Recycle solvent
Composition (kg/sec)										
H ₂	--	22.85	--	--	11.44	9.35	9.35	--	--	--
C ₁ -C ₄	--	24.83	--	--	12.44	22.66	22.66	--	--	--
C ₅ -465°K	--	--	--	--	--	9.75	--	9.75	--	--
465°K-650°K	--	--	--	--	--	} 486.06	--	--	} 400.46	--
650°K-810°K	--	--	346.41	--	--		--	--		85.60
>810°K	--	--	--	--	--		--	--		--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	186.71	--	--	--	--	--	--	--	--	--
H ₂ O	--	0.23	--	9.68	0.12	} 19.62	} 19.62	--	--	--
H ₂ S	--	--	--	--	--			--	--	--
NH ₃	--	--	--	--	--			--	--	--
CO ₂	--	--	--	--	--			--	--	--
Air	--	--	--	--	--	--	--	--	--	--
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	186.71	47.91	346.41	9.68	24.00	547.44	51.63	9.75	400.46	85.60

TABLE 5-5 (continued)

Stream No.	11	12	13	14	15	16	17	18	19	20
Component Description	Hydrogen to Solution Hydro-cracking	Solution Hydro-cracker Effluent	Hydro-cracking Product Gas	Distillate	Heavy Coal Liquids	Recycle Solvent	Acid Gases	Coal Liquids	Gas-Make to Fuel	Combustion Air
Composition (kg/sec)										
H ₂	11.41	12.18	12.18	--	--	--	--	--	} 0.68	--
C ₁ -C ₄	12.39	54.65	54.65	--	--	--	--	--		--
C ₅ -465°K	--	} 77.50	--	} 77.50	--	--	--	--	--	--
465°K-650°K	--		--		--	--	--	--	--	--
650°K-810°K	--	} 316.13	--	--	} 55.32	260.81	--	18.31	--	--
>810°K	--		--	--		--	--	--	--	--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	--	--	--	--	--	--	--	--	--	--
H ₂ O	0.11	} 25.29	} 8.06	--	--	--	} 17.23	--	--	--
H ₂ S	--			--	--	--		--	--	--
NH ₃	--			--	--	--		--	--	--
CO ₂	--			--	--	--		--	--	--
Air	--	--	--	--	--	--	--	--	--	19.67
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	23.91	485.75	74.89	77.50	55.32	260.81	17.23	18.31	0.68	19.67

TABLE 5-5 (continued)

Stream No.	21	22	23	24	25	26	27	28	29	29	
Component Description	Flue Gas	Refuse	Hydro-gasifier Effluent	Carbon Dioxide	Sulfur	Ammonia	Waste-water	Product Aromatics	Reformer Feed Gas	Reformer Feed Composition (kg-mole/sec)	
Composition (kg/sec)											
H ₂	--	--	2.20	--	--	--	--	--	2.20	H ₂	1.093
C ₁ -C ₄	--	--	138.85	--	--	--	--	--	138.85	CH ₄	5.735
C ₅ -465°K	--	--	18.93	--	--	--	--	18.93	--	C ₂ H ₆	1.126
465°K-650°K	--	--	--	--	--	--	--	--	--	C ₃ H ₈	0.147
650°K-810°K	--	--	--	--	--	--	--	--	--	C ₄ H ₁₀	0.114
>810°K	--	--	--	--	--	--	--	--	--	Total	8.215
Coke (inc ash)	--	34.11	--	--	--	--	--	--	--	(kg-mole/sec)	
Coal	--	--	--	--	--	--	--	--	--	m ³ /sec (STP)	184.1
H ₂ O	--	--	10.72	--	--	--	17.19	--	--		
H ₂ S	--	--		--	4.76	--	--	--	--		
NH ₃	--	--		--	--	2.89	--	--	--		
CO ₂	--	--		3.11	--	--	--	--	--		
Air	--	--	--	--	--	--	--	--	--		
Flue Gas	21.89	--	--	--	--	--	--	--	--		
Total kg/sec	21.89	34.11	170.70	3.11	4.76	2.89	17.19	18.93	141.05		

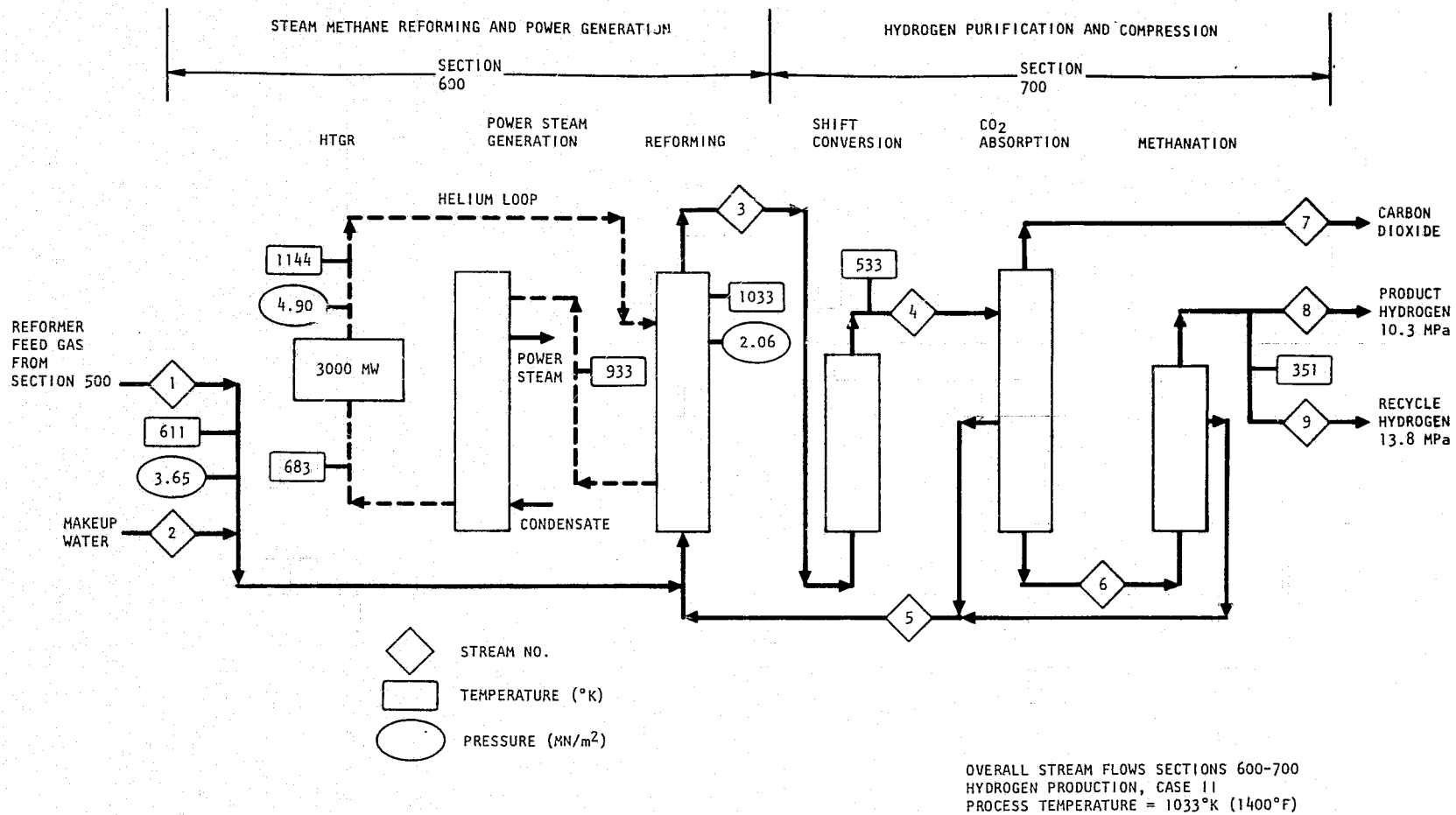
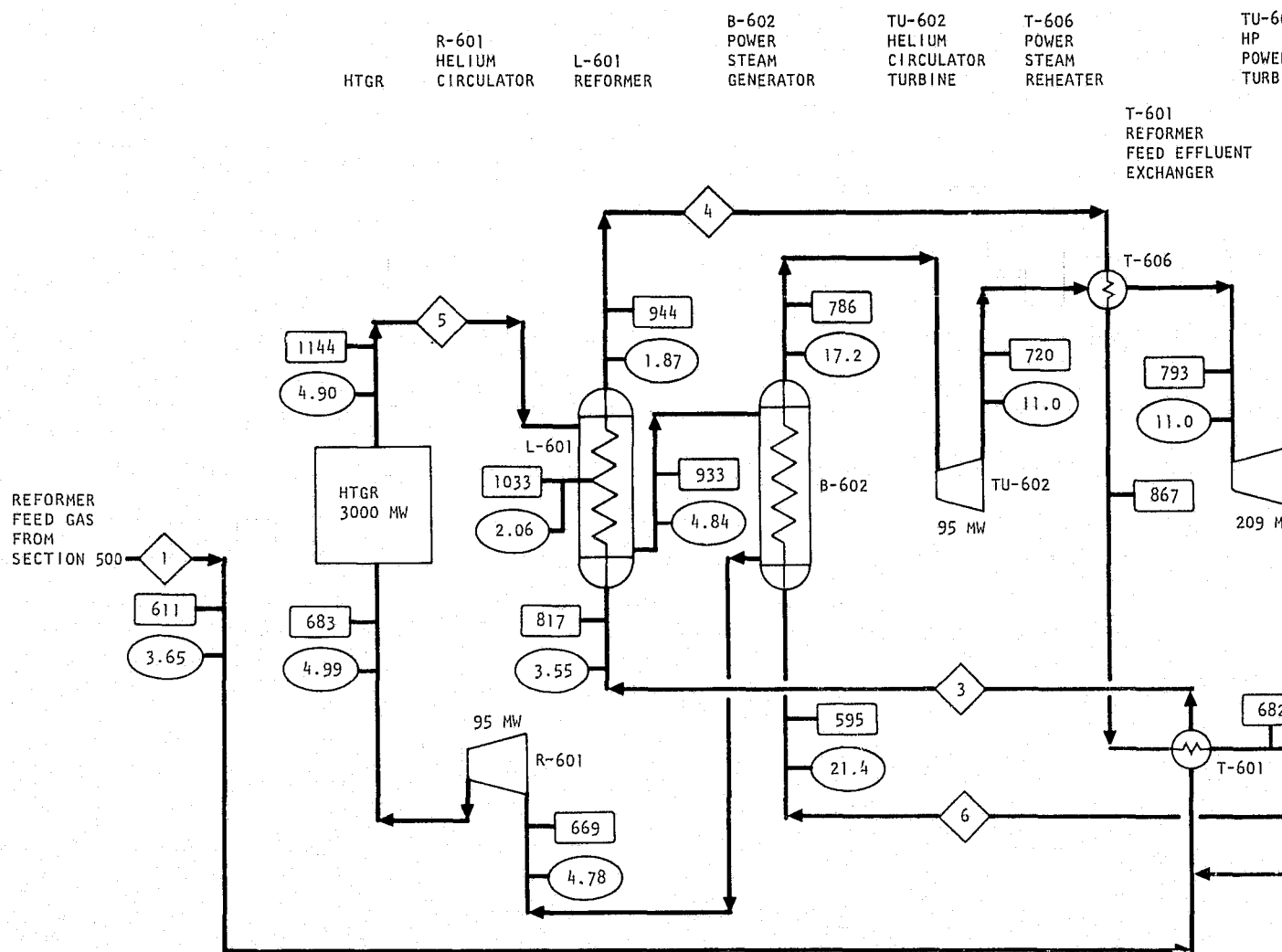


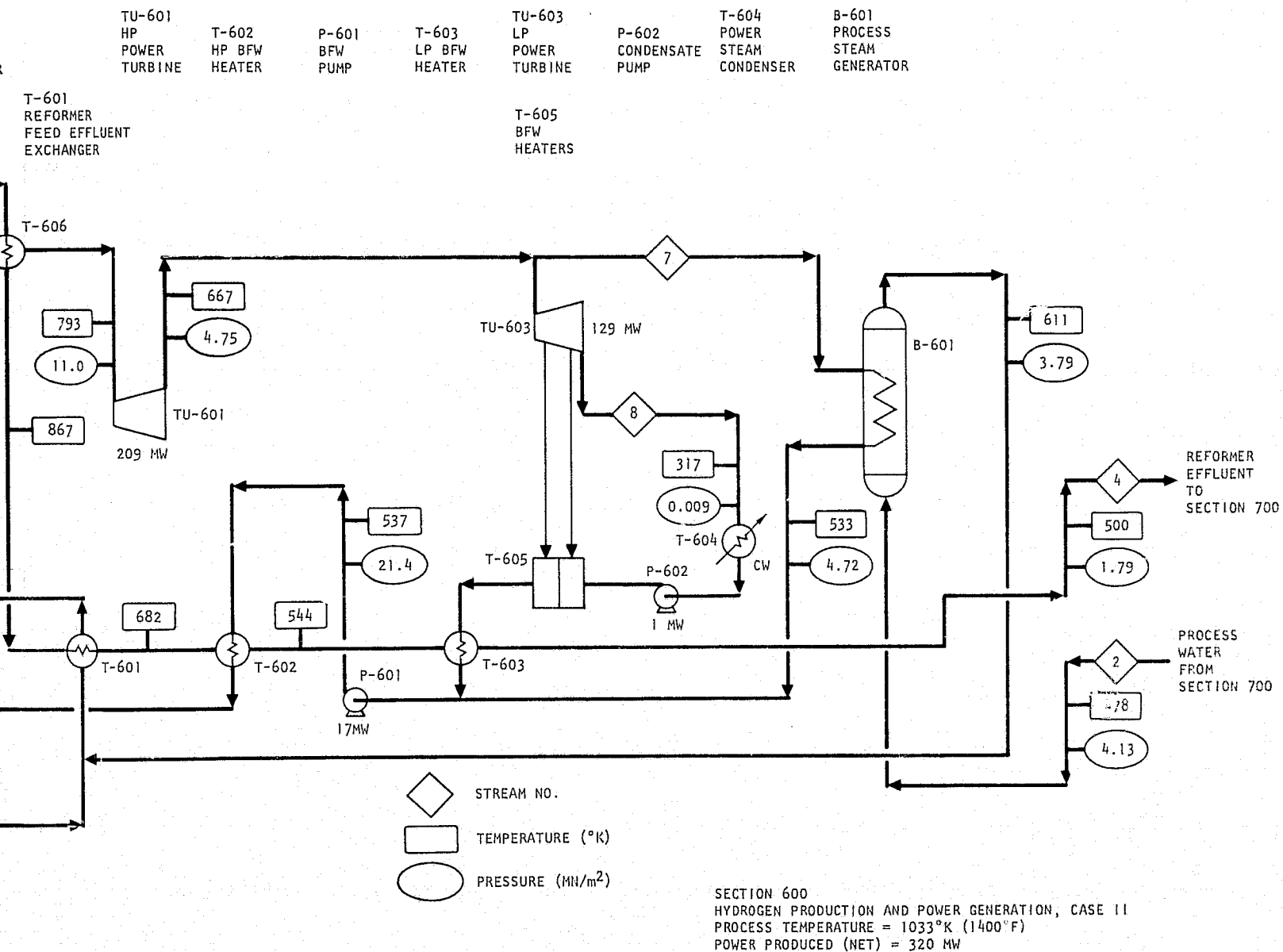
Fig. 5-2. Flow diagram, Sections 600-700, Case II

TABLE 5-6
HYDROGEN PRODUCTION AND POWER GENERATION OVERALL STREAM FLOWS, CASE II, PROCESS TEMPERATURE = 1033°K (1400°F)

Stream No.	1	2	3	4	5	6	7	8	9	Stream No.	8	9
Component Description	Reformer Feed Gas	Makeup Water	Reformer Effluent	Shift Converter Effluent	Recycle Condensate	Carbon Dioxide Absorber Effluent	Carbon Dioxide	Recycle Hydrogen to Coal Liquefaction	Product Hydrogen	Hydrogen Composition (kg/sec)	Recycle Hydrogen	Product Hydrogen
Composition (kg-mole/sec)												
H ₂	1.093	--	21.245	23.259	--	23.186	0.073	11.346	11.156	H ₂	22.85	22.47
CH ₄	5.735	--	2.875	2.875	--	2.875	--	1.549	1.524	CH ₄	24.83	24.42
C ₂ H ₆	1.126	--	--	--	--	--	--	--	--	CO	--	--
C ₃ H ₈	0.147	--	--	--	--	--	--	--	--	CO ₂	--	--
C ₄ H ₁₀	0.114	--	--	--	--	--	--	--	--	H ₂ O	0.23	0.22
CO	--	--	2.122	0.108	--	0.108	--	--	--	Total kg/sec	47.91	47.11
CO ₂	--	--	3.887	5.901	--	0.090	5.811	--	--			
H ₂ O	--	12.064	30.187	28.173	28.019	2.194	0.417	0.013	0.012			
Total kg-mole/sec	8.215	12.064	60.316	60.316	28.019	28.453	6.301	12.908	12.692			
kg/sec	141.05	217.14	862.50	862.50	504.31	139.24	263.17	47.91	47.11			



FOLDOUT FRAME



FOLDOUT FRAME 2

Fig. 5-3. Flow diagram, Section 600, Case II

TABLE 5-7

STEAM METHANE REFORMING AND POWER GENERATION STREAM FLOWS, CASE II, PROCESS TEMPERATURE = 1033°K (1400°F)

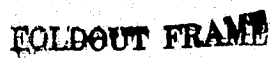
Stream No.	1	2	3	4	5	6	7	8
Component Description	Reformer Feed Gas	Reformer Feed-water	Reformer Feed	Reformer Effluent	Helium Heat Transfer Loop	Boiler Feed-water	Steam to Process Steam Generator	Steam to LP Turbine
Composition (kg-mole/sec)								
He	--	--	--	--	--	--	--	--
H ₂	1.093	--	1.093	21.245	312.372	--	--	--
CH ₄	5.735	--	5.735	2.875	--	--	--	--
C ₂ H ₆	1.126	--	1.126	--	--	--	--	--
C ₃ H ₈	0.147	--	0.147	--	--	--	--	--
C ₄ H ₁₀	0.114	--	0.114	--	--	--	--	--
CO	--	--	--	2.122	--	--	--	--
CO ₂	--	--	--	3.887	--	--	--	--
H ₂ O	--	40.083	40.083	30.187	--	51.612	42.948	7.337
Total kg-mole/sec	8.215	40.083	48.298	60.316	312.372	51.612	42.948	7.337
Total kg/sec	141.05	721.45	862.50	862.50	1251.46	928.96	773.02	132.06

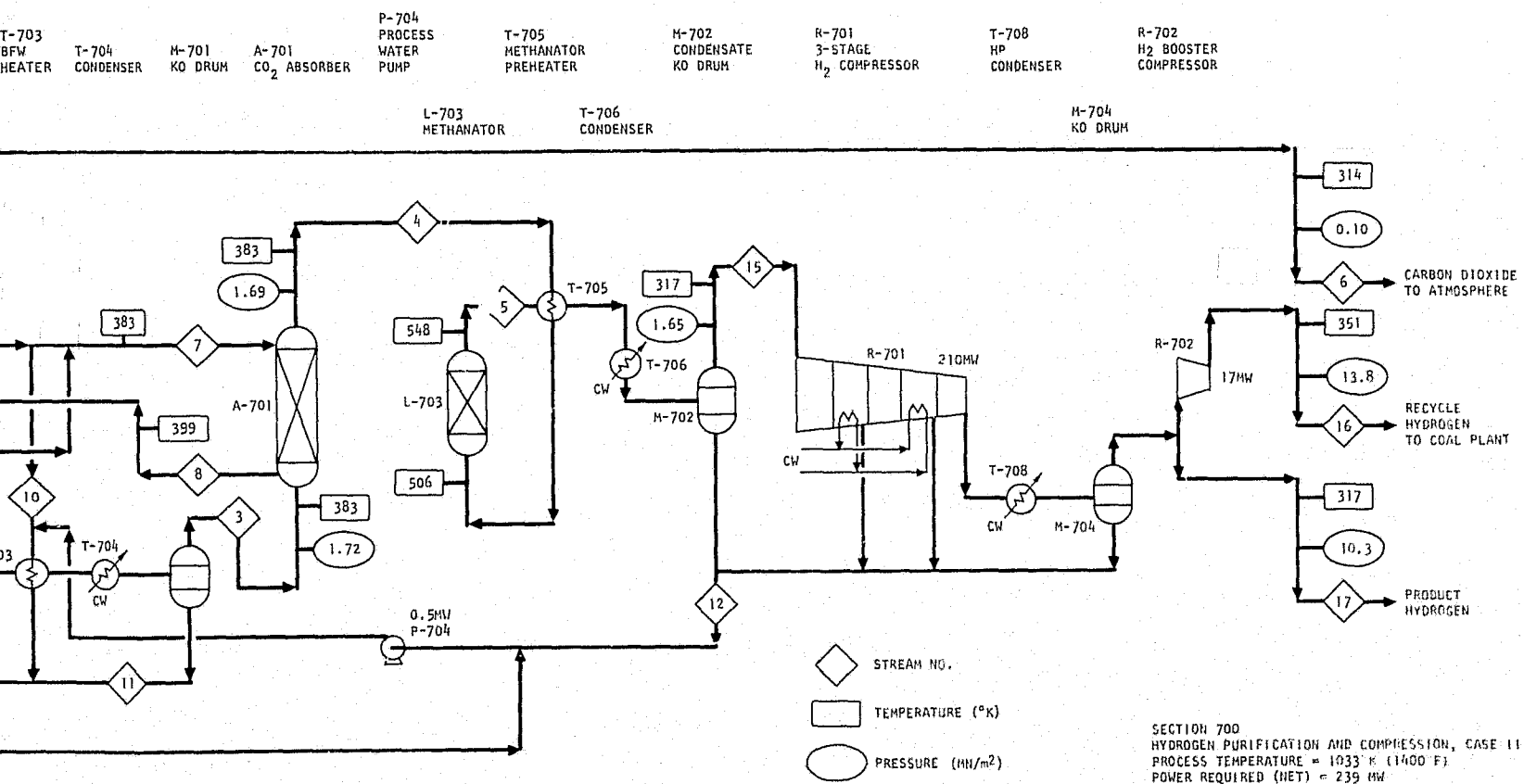
	<u>Thermal Power (MW)</u>
Reformer	1377
Steam generator	<u>1718</u>
Subtotal	3095
Helium circulators	<u>-95</u>
Total HTGR power	3000

Reformer feed gas from the hydrogasification product separation (Section 500) is mixed with steam and preheated to 817°K (1010°F). An integral return tube reformer design is employed (Ref. 5-1). The reformer feed passes over catalyst beds in the annulus of the reformer tubes. The process gas leaving the catalyst beds at 1033°K (1400°F) and 2.06 MN/m² (300 psia) exits through integral return tubes where a portion of the sensible heat contained in the reformer effluent gas is recovered. The process gas exits at 944°K (1240°F) and is further cooled to 500°K (440°F) by heat exchange with steam (exchanger T-606), reformer feed (exchanger T-601), and boiler feedwater (exchangers T-602 and T-603). The reformer effluent is then sent forward to Section 700 for shift conversion and further purification.

Steam is generated by the HTGR steam generators at 786°K (955°F) and 17.2 MN/m² (2500 psia). The high-pressure steam is used to provide power for helium circulation and is then reheated to provide additional process power, primarily for hydrogen compression. The majority of the steam is condensed in the process steam generators (B-601), providing steam required for the reformer feed. The remaining steam is sent to condensing turbines, providing additional power. Condensate is recycled as boiler feedwater after it is preheated by the reformer effluent in exchangers T-602 and T-603.

5.2.1.2. Hydrogen Purification and Compression (Section 700). The following description refers to the process flow sheet shown in Fig. 5-4 and the mass balance given in Table 5-8.





FOLDOUT FRAME 2

Fig. 5-4. Flow diagram, Section 700, Case II

TABLE 5-8
HYDROGEN PURIFICATION AND COMPRESSION STREAM FLOWS, CASE II, PROCESS TEMPERATURE = 1033°K (1400°F)

Stream No.	1	2	3	4	5	6	7	8
Component Description	Reformer Effluent	Shift Converter Effluent	Carbon Dioxide Absorber Feed	Carbon Dioxide Absorber Effluent	Methanator Effluent	Carbon Dioxide to Atmosphere	Lean Solution to Absorber	Rich Solution to Stripper
Composition (kg-mole/sec)								
H ₂	21.245	23.259	23.259	23.186	22.502	0.073	0.004	0.077
CH ₄	2.875	2.875	2.875	2.875	3.073	--	--	--
CO	2.122	0.108	0.108	0.108	--	--	--	--
CO ₂	3.887	5.901	5.901	0.090	--	5.811	--	--
H ₂ O	30.187	28.173	2.916	2.194	2.482	0.417	109.931	104.842
K ₂ CO ₃	--	--	--	--	--	--	8.716	2.905
KHCO ₃	--	--	--	--	--	--	1.938	13.560
As ₂ O ₃	--	--	--	--	--	--	0.248	0.248
Total kg-mole/sec	60.316	60.316	35.059	28.453	28.057	6.301	120.837	121.632
kg/sec	862.50	862.50	407.90	139.24	139.24	263.17	3418.84	3687.50

TABLE 5-8 (continued)

Stream No.	9	10	11	12	13	14	15	16	17
Component Description	Lean Solution From Stripper	Recycle Condensate	Recycle Condensate	Recycle Condensate	Makeup Water	Reformer Feed-water	Hydrogen Compressor Feed	Hydrogen Recycle to Coal Plant	Product Hydrogen
Composition (kg-mole/sec)									
H ₂	0.004	--	--	--	--	--	22.502	11.346	11.156
CH ₄	--	--	--	--	--	--	3.073	1.549	1.524
CO	--	--	--	--	--	--	--	--	--
CO ₂	--	--	--	--	--	--	--	--	--
H ₂ O	94.207	0.305	25.257	2.457	12.064	40.083	0.140	0.013	0.012
K ₂ CO ₃	8.716	--	--	--	--	--	--	--	--
KHCO ₃	1.938	--	--	--	--	--	--	--	--
As ₂ O ₃	0.248	--	--	--	--	--	--	--	--
Total kg-mole/sec	105.113	0.305	25.257	2.457	12.064	40.083	25.715	12.908	12.692
kg/sec	3135.82	5.49	454.60	44.22	217.14	721.45	97.09	47.91	47.11

Reformer effluent at 500°C (440°F) is fed to the shift converters (L-702) for further conversion of CO by the water gas shift reaction. Shift converter effluent at 533°K (500°F) is cooled by heat exchange with process boiler feedwater in exchanger T-702, and excess steam in the process gas is condensed in the CO₂ stripper reboiler (T-701), the BFW heaters (T-703), and the condenser (T-704). Condensate is removed in the high-temperature knockout drum (M-701) and recycled for treatment. The process gas is scrubbed in the CO₂ absorber (A-701) by aqueous hot potassium carbonate to remove CO₂. Remaining carbon oxide impurities are reduced to trace levels in the methanator (L-703). The methanator effluent is cooled in the methanator preheater (T-705) and condenser T-708. Condensate is recycled and the hydrogen stream is sent forward for compression.

The rich hot carbonate solution is regenerated in the CO₂ stripper (A-702), and the lean solution is returned to the absorber by the lean solution circulation pump (P-701). The CO₂ stripper effluent gas is cooled in the reflux condenser (T-707), and the condensate is recycled. Carbon dioxide is vented to the atmosphere.

Hydrogen is compressed to product delivery pressure [10.34 MN/m² (1500 psia)] in multistage centrifugal compressors (R-701). A portion of the hydrogen stream is compressed to 13.78 MN/m² (2000 psia) by the hydrogen booster compressor (R-702) and then recycled for hydrogasification of coal. The hydrogen product has a purity of 87.9 mol %.

5.2.2. Case I, Process Temperature = 922°K (1200°F)

A summary of the plant performance is given in Table 5-9, and additional details on the process power requirements and cooling water requirements are given in Tables 5-10 and 5-11. Overall stream flows are given in Figs. 5-5 and 5-6 and Tables 5-12 and 5-13.

The principal modification to the base case [process temperature = 1033°K (1400°F)] was the addition of a cryogenic separation step to improve

TABLE 5-9
OVERALL PERFORMANCE SUMMARY, CASE I, PROCESS TEMPERATURE = 922°K (1200°F)

	Quantity at 100% Capacity	Gross Heat Value		% of Total Input
		(MW)	(10 ⁹ Btu/hr)	
Input				
Coal feed	98.66 kg/sec (9396 tpd)	2859	9.76	48.8
Makeup water	127.88 kg/sec (2028 gpm)			
HTGR	3000 MW	3000	10.24	51.2
Total in		5859	20.00	100.0
Product				
Hydrogen	7.972 kg-mole/sec (545 MMscfd) 95.4 mol % ^(a)	2498	8.53	42.7
Aromatics	10.19 kg/sec (5550 bpd)	431	1.47	7.4
Power	138 MW	138	0.47	2.3
Subtotal		3067	10.47	52.4
Waste Heat				
Char	18.01 kg/sec (1715 tpd)	337	1.15	5.7
To cooling water		2338	7.98	39.9
Other		117	0.40	2.0
Subtotal		2792	9.53	47.6
Total out		5859	20.00	100.0

^(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

TABLE 5-10
POWER REQUIREMENTS, CASE I,
PROCESS TEMPERATURE = 922°K (1200°F)

	Power (MW)
Section 600	
Power produced	
Circulator turbine	101
High-pressure turbine	209
Low-pressure turbine	165
Subtotal	475
Power consumed	
Helium circulator	(101)
Pumps	(25)
Subtotal	(126)
Net power produced, Section 600	349
Section 700	
Power produced	--
Power consumed	
Hydrogen compression	(143)
Pumps	(9)
Other	(46)
Subtotal	(198)
Net power consumed, Section 700	(198)
Net power consumed (Sections 100 to 500)	(13)
Total net power produced (Sections 100 to 700)	138

TABLE 5-11
COOLING WATER REQUIREMENTS, CASE I,
PROCESS TEMPERATURE = 922°K (1200°F)

Item No.	Description	Heat Duty	
		MW	10 ⁹ Btu/hr
Sections 600, 700			
T-604	Power steam condenser	669	2.284
T-704	Condenser	419	1.431
T-706	Condenser	141	0.483
T-707	CO ₂ stripper condenser	647	2.211
T-708	Condenser	59	0.201
R-701	Compressor intercoolers	60	0.202
Total Sections 600 to 700		1995 MW	(6812 MMBtu/hr)
Total Sections 100 to 500		343 MW	(1170 MMBtu/hr)
Total heat duty		2338 MW	(7982 MMBtu/hr)

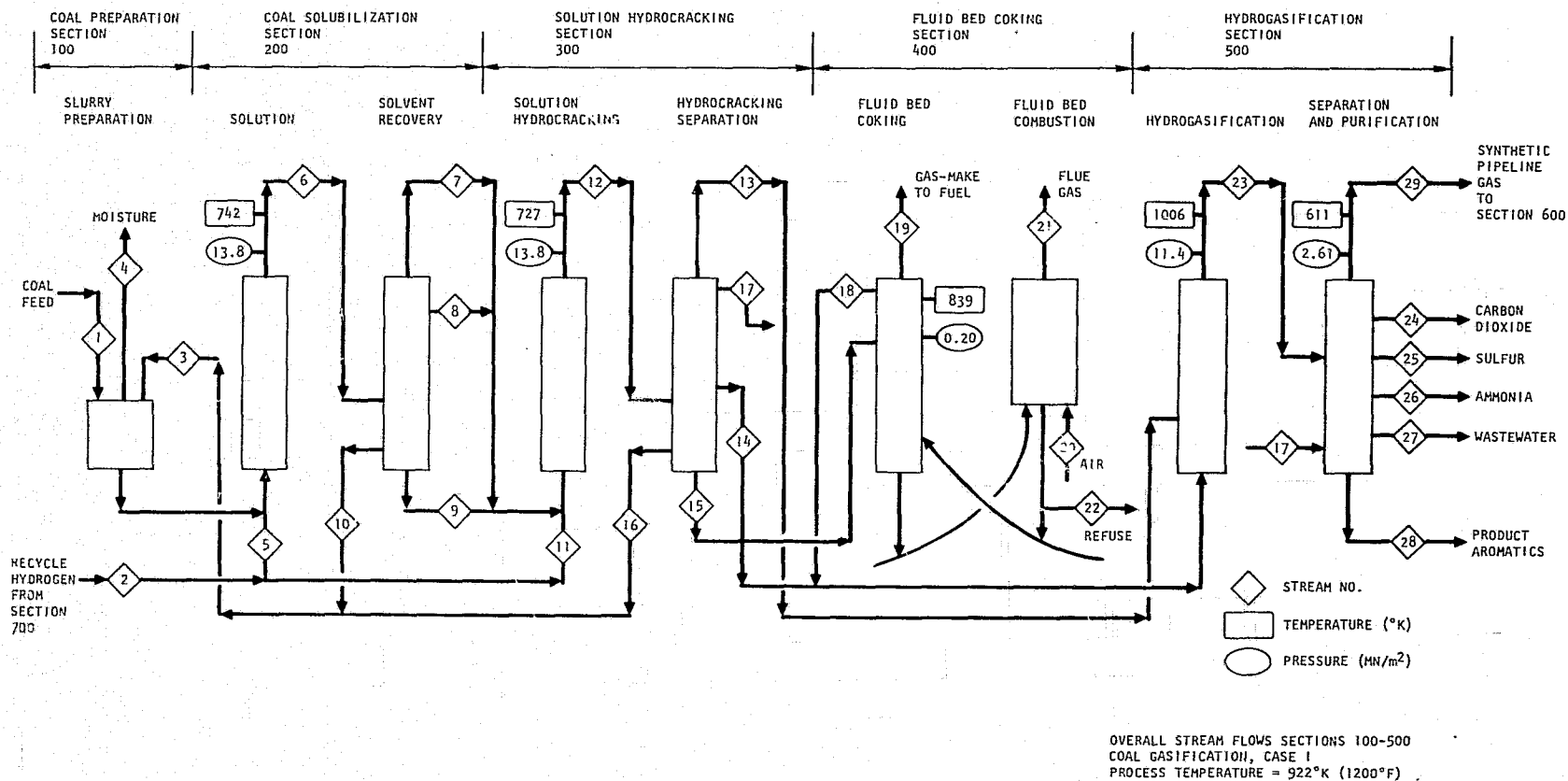


Fig. 5-5. Flow diagram, Sections 100-500, Case I

TABLE 5-12
HYDROGASIFICATION OF COAL LIQUIDS OVERALL PLANT FLOWS, CASE I, PROCESS TEMPERATURE = 922°K (1200°F)

Stream No.	1	2	3	4	5	6	7	8	9	10
Component Description	Coal Feed	Recycle Hydrogen	Recycle Solvent	Moisture	Hydrogen to Coal Solution	Coal Solution Effluent	Hydrocarbon Gases	Distillate	Coal Liquids	Recycle Solvent
Composition (kg/sec)										
H ₂	--	12.08	--	--	6.05	4.42	4.42	--	--	--
C ₁ -C ₄	--	4.60	--	--	2.30	8.28	8.28	--	--	--
C ₅ -465°K	--	--	--	--	--	5.15	--	5.15	--	--
465°K-650°K	--	--	--	--	--	256.84	--	--	211.61	--
650°K-810°K	--	--	183.05	--	--		--	--		45.23
>810°K	--	--	--	--	--		--	--		--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	98.66	--	--	--	--	--	--	--	--	--
H ₂ O	--	--	--	5.12	--	10.25	10.25	--	--	--
H ₂ S	--	--	--	--	--			--	--	--
NH ₃	--	--	--	--	--			--	--	--
CO ₂	--	--	--	--	--			--	--	--
Air	--	--	--	--	--	--	--	--	--	--
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	98.66	16.68	183.05	5.12	8.35	284.94	22.95	5.15	211.61	45.23

TABLE 5-12 (continued)

Stream No.	11	12	13	14	15	16	17	18	19	20
Component Description	Hydrogen to Solution Hydro-cracking	Solution Hydro-cracker Effluent	Hydro-cracking Product Gas	Distillate	Heavy Coal Liquids	Recycle Solvent	Acid Gases	Coal Liquids	Gas-Make to Fuel	Combustion Air
Composition (kg/sec)										
H ₂	6.03	6.13	6.13	--	--	--	--	--	} 0.36	--
C ₁ -C ₄	2.30	20.67	20.67	--	--	--	--	--		--
C ₅ -465°K	--	} 40.95	--	} 40.95	--	--	--	--		--
465°K-650°K	--		--		--	--	--	--	--	--
650°K-810°K	--	} 167.05	--	--	} 29.23	137.82	--	9.68	--	--
>810°K	--		--	--		--	--	--	--	--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	--	--	--	--	--	--	--	--	--	--
H ₂ O	--	} 13.24	} 4.22	--	--	} 9.02	--	--	--	--
H ₂ S	--			--	--		--	--	--	--
NH ₃	--			--	--		--	--	--	--
CO ₂	--			--	--		--	--	--	--
Air	--	--	--	--	--	--	--	--	--	10.39
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	8.33	248.04	31.02	40.95	29.23	137.82	9.02	9.68	0.36	10.39

TABLE 5-12 (continued)

Stream No.	21	22	23	24	25	26	27	28	29	29	
Component Description	Flue Gas	Refuse	Hydro-gasifier Effluent	Carbon Dioxide	Sulfur	Ammonia	Waste-water	Product Aromatics	Reformer Feed Gas	Reformer Feed Composition (kg-mole/sec)	
Composition (kg/sec)											
H ₂	--	--	1.16	--	--	--	--	--	1.16	H ₂	0.577
C ₁ -C ₄	--	--	64.67	--	--	--	--	--	64.67	CH ₄	2.488
C ₅ -465°K	--	--	10.19	--	--	--	--	10.19	--	C ₂ H ₆	0.595
465°K-650°K	--	--	--	--	--	--	--	--	--	C ₃ H ₈	0.078
650°K-810°K	--	--	--	--	--	--	--	--	--	C ₄ H ₁₀	0.060
>810°K	--	--	--	--	--	--	--	--	--	Total	3.798
Coke (inc ash)	--	18.01	--	--	--	--	--	--	--	(kg-mole/sec)	
Coal	--	--	--	--	--	--	--	--	--		
H ₂ O	--	--	5.63	--	--	--	8.96	--	--		
H ₂ S	--	--		--	2.52	--	--	--	--		
NH ₃	--	--		--	--	1.53	--	--	--		
CO ₂	--	--		1.64	--	--	--	--	--		
Air	--	--	--	--	--	--	--	--	--		
Flue Gas	11.57	--	--	--	--	--	--	--	--		
Total kg/sec	11.57	18.01	81.65	1.64	2.52	1.53	8.96	10.19	65.83		

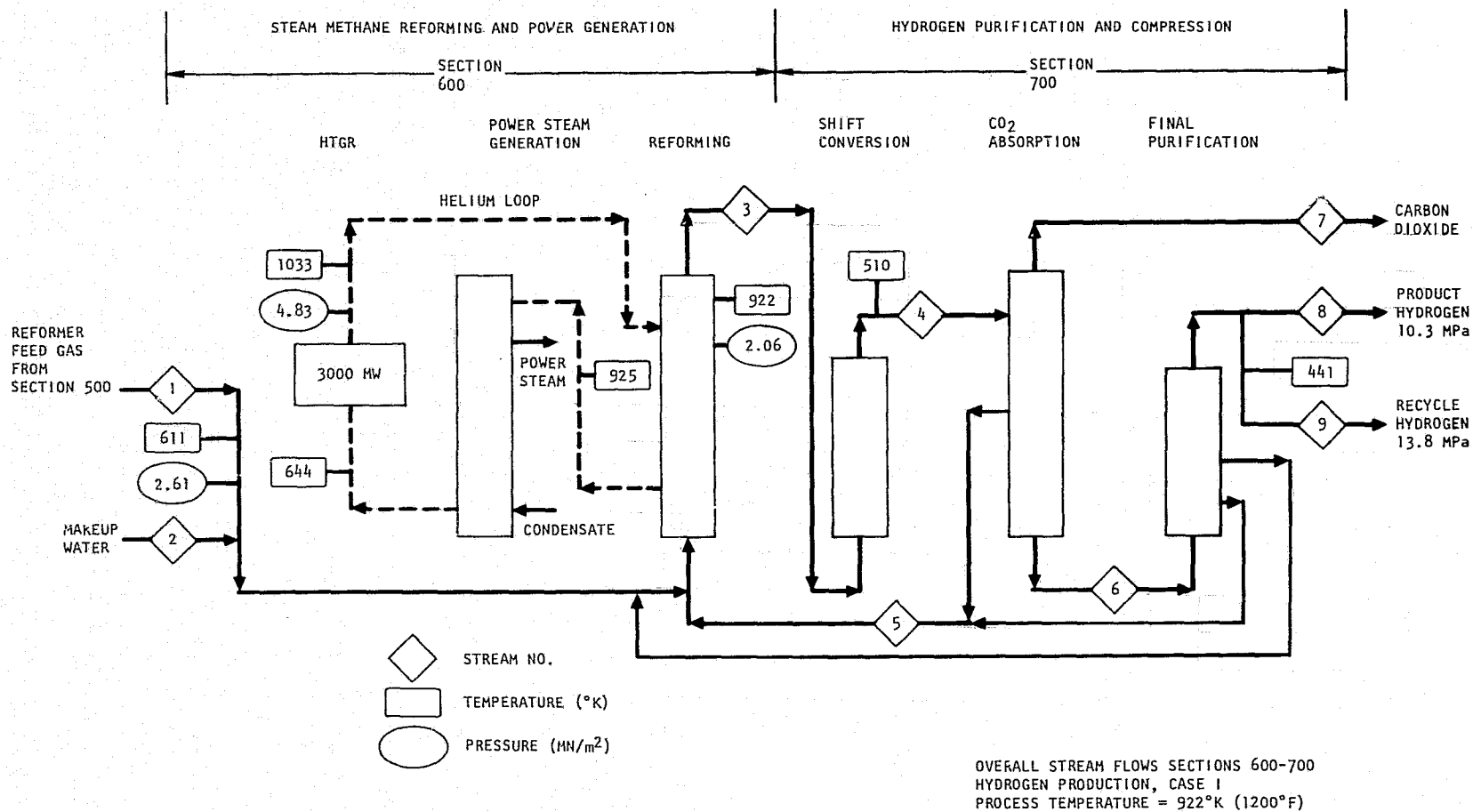


Fig. 5-6. Flow diagram, Sections 600-700, Case I

TABLE 5-13

HYDROGEN PRODUCTION AND POWER GENERATION OVERALL STREAM FLOWS, CASE I, PROCESS TEMPERATURE = 922°K (1200°F)

Stream No.	1	2	3	4	5	6	7	8	9	Stream No.	8	9
Component Description	Reformer Feed Gas	Makeup Water	Reformer Effluent	Shift Converter Effluent	Recycle Condensate	Carbon Dioxide Absorber Effluent	Carbon Dioxide	Recycle Hydrogen to Coal Liquefaction	Product Hydrogen	Hydrogen Composition (kg/sec)	Recycle Hydrogen	Product Hydrogen
Composition (kg-mole/sec)												
H ₂	0.577	--	13.640	14.242	--	14.195	0.047	5.998	7.609	H ₂	12.08	15.32
CH ₄	2.488	--	5.562	5.562	--	5.562	--	0.287	0.363	CH ₄	4.60	5.82
C ₂ H ₆	0.595	--	--	--	--	--	--	--	--	CO	--	--
C ₃ H ₈	0.078	--	--	--	--	--	--	--	--	CO ₂	--	--
C ₄ H ₁₀	0.060	--	--	--	--	--	--	--	--	H ₂ O	--	--
CO	--	--	0.623	0.021	--	0.021	--	--	--	Total kg/sec	16.68	21.14
CO ₂	--	--	2.976	3.578	--	0.076	3.502	--	--			
H ₂ O	--	7.105	34.645	34.043	34.115	1.748	0.101	--	--			
Total kg-mole/sec	3.798	7.105	57.446	57.446	34.115	21.602	3.650	6.285	7.972			
kg/sec	65.83	127.88	888.48	888.48	614.04	153.13	155.89	16.68	21.14			

the purity of the hydrogen product and hydrogen recycle streams and to conserve feedstock. At the low reforming temperatures considered in this case, reformer equilibrium restrictions severely limit methane conversion, and undesirably large quantities of methane would be present in the product hydrogen if additional purification is not provided. In addition, the pressure in the coal solution and hydrogasification reactors would have to be substantially increased to maintain the partial pressures of hydrogen required in these reactors.

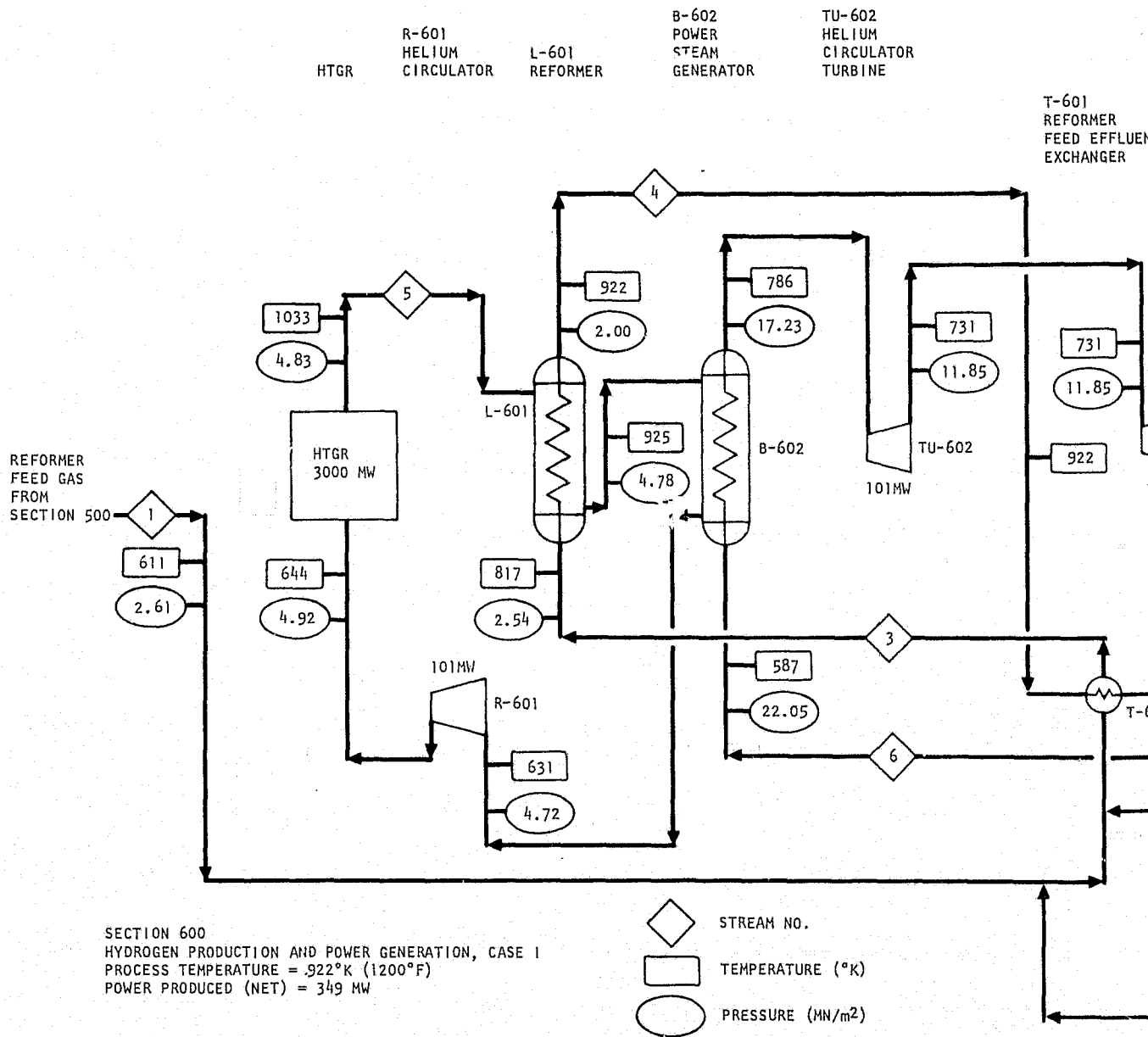
5.2.2.1. Hydrogen Production and Power Generation (Section 600). The detailed changes in process conditions from the base case are given in the process flow sheet shown in Fig. 5-7 and the mass balance in Table 5-14.

Hot helium at 1033°K (1400°F) transfers heat from the HTGR nuclear core to the reformers and steam generators. The distribution of HTGR thermal power is as follows:

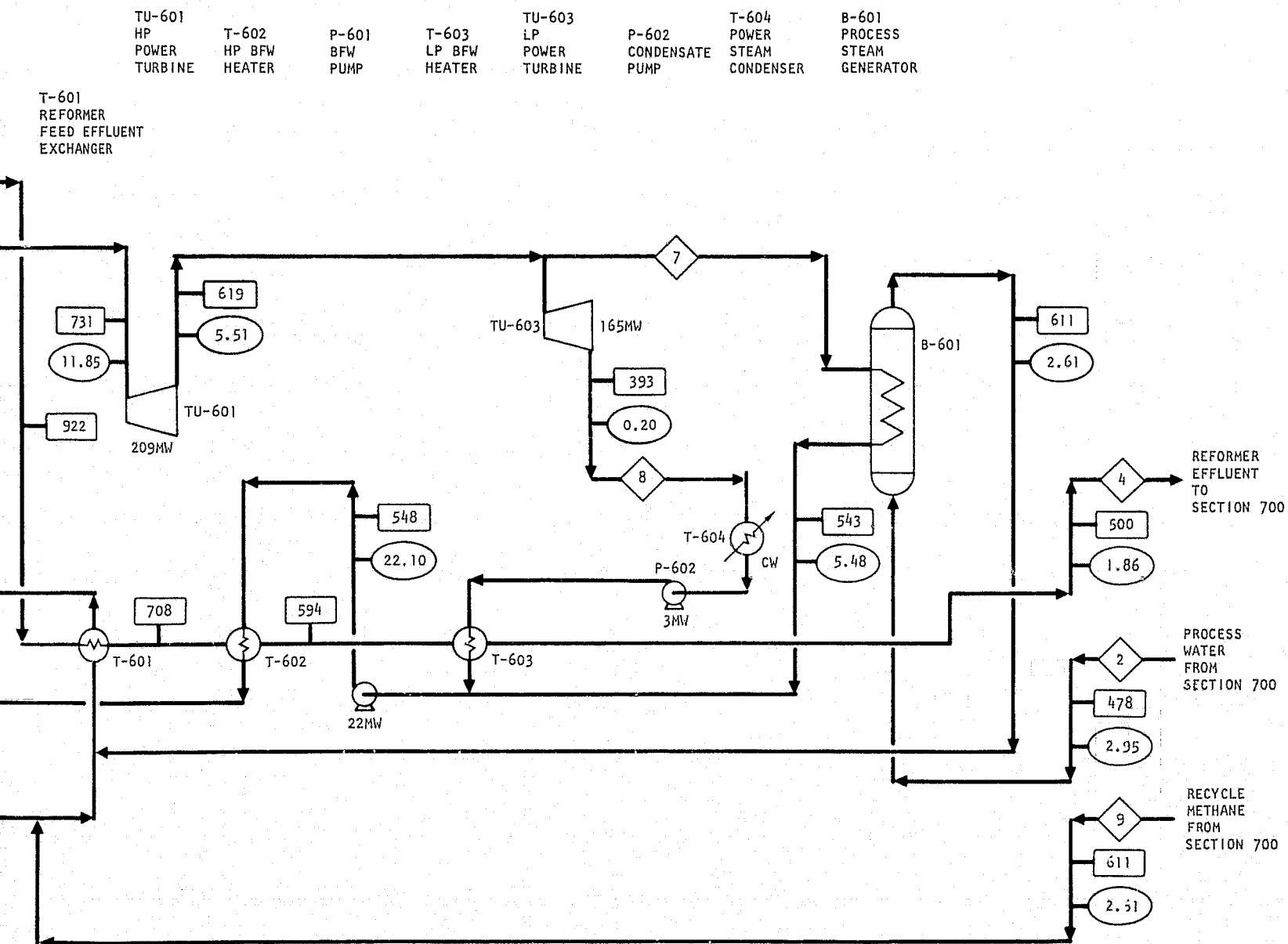
	<u>Thermal Power (MW)</u>
Reformers	849
Steam generators	<u>2240</u>
Subtotal	3089
Helium circulators	<u>-89</u>
Total HTGR power	3000

A common return tube reformer design is employed in this case, with the effluent gas leaving the reformer at essentially catalyst bed discharge conditions of 922°K (1200°F) and 2.06 MN/m² (300 psia). The low temperature of the reformer effluent gas requires the elimination of the power steam reheater. All other process steps are identical to the base case.

5.2.2.2. Hydrogen Purification and Compression (Section 700). The detailed changes in process conditions from the base case are given in the process flow sheet shown in Fig. 5-8 and the mass balance given in Table 5-15.



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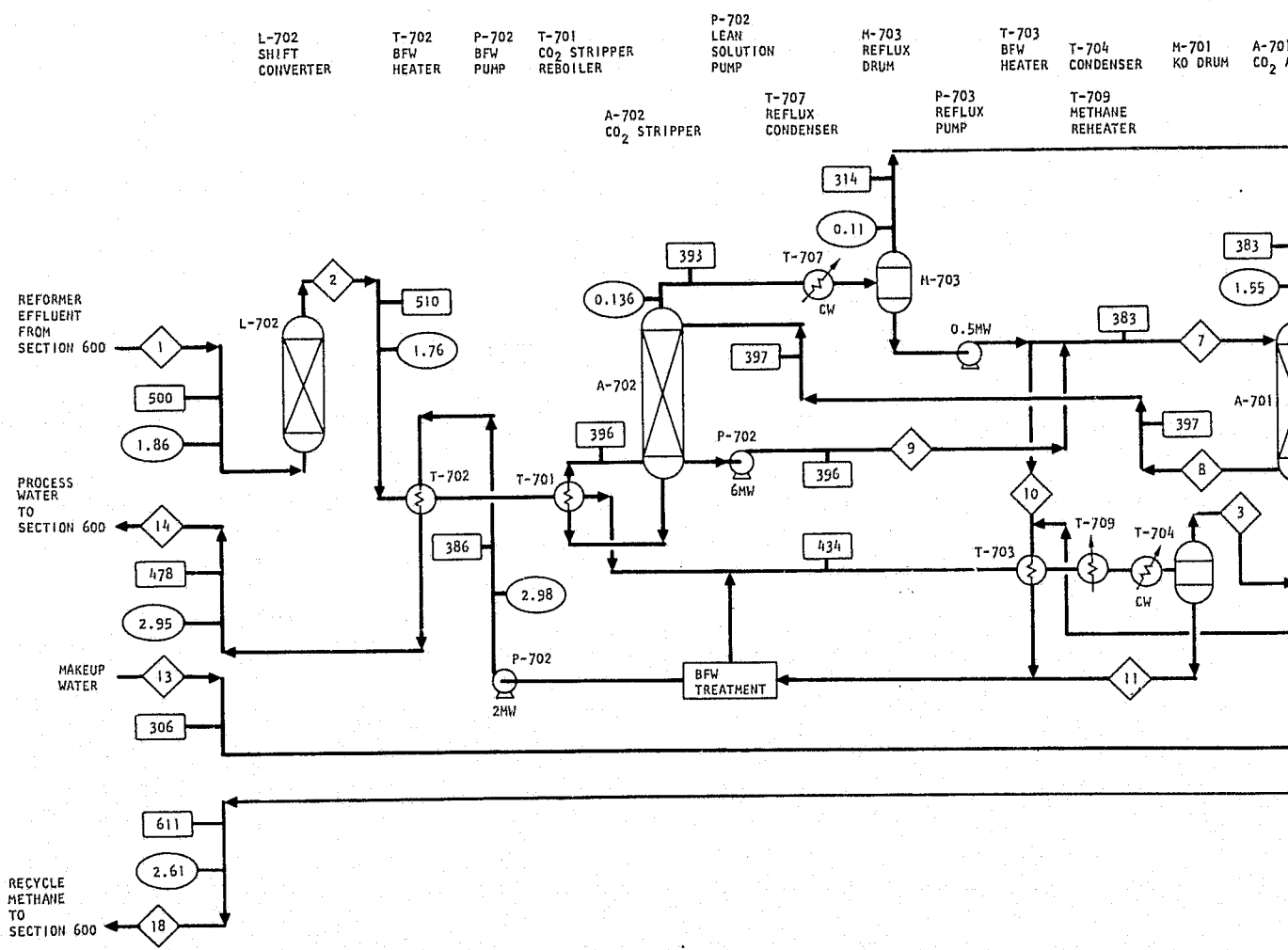


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Fig. 5-7. Flow diagram, Section 600, Case I

TABLE 5-14
HYDROGEN PRODUCTION AND POWER GENERATION STREAM FLOWS, CASE I, PROCESS TEMPERATURE = 922°K (1200°F)

Stream No.	1	2	3	4	5	6	7	8	9
Component Description	Reformer Feed Gas	Reformer Feed-water	Reformer Feed	Reformer Effluent	Helium Heat Transfer Loop	Power Cycle BFW	Steam to Process Steam Generator	Steam From LP Turbine	Recycle Methane
Composition (kg-mole/sec)									
He	--	--	--	--	369.18	--	--	--	--
H ₂	0.577	--	0.798	13.640	--	--	--	--	0.221
CH ₄	2.488	--	7.497	5.562	--	--	--	--	5.009
C ₂ H ₆	0.595	--	0.595	--	--	--	--	--	--
C ₃ H ₈	0.078	--	0.078	--	--	--	--	--	--
C ₄ H ₁₀	0.060	--	0.060	--	--	--	--	--	--
CO	--	--	--	0.623	--	--	--	--	--
CO ₂	--	--	--	2.976	--	--	--	--	--
H ₂ O.	--	41.220	41.220	34.645	--	65.552	48.653	16.899	--
Total kg-mole/sec	3.798	41.220	50.248	57.446	369.18	65.552	48.653	16.899	5.230
Total kg/sec	65.83	741.92	888.48	888.48	1476.72	1179.87	875.71	304.16	80.73



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TABLE 5-15
HYDROGEN PURIFICATION AND COMPRESSION STREAM FLOWS, CASE I, PROCESS TEMPERATURE = 922°K (1200°F)

Stream No.	1	2	3	4	5	6	7	8	9
Component Description	Reformer Effluent	Shift Converter Effluent	CO ₂ Absorber Feed	CO ₂ Absorber Effluent	Methanator Effluent	CO ₂ to Atmosphere	Lean Solution to Absorber	Rich Solution to Stripper	Lean Solution From Stripper
Composition (kg-mole/sec)									
H ₂	13.640	14.242	14.242	14.195	13.828	0.047	0.003	0.050	0.003
CH ₄	5.562	5.562	5.562	5.562	5.659	--	--	--	--
CO	0.623	0.021	0.021	0.021	--	--	--	--	--
CO ₂	2.976	3.578	3.578	0.076	--	3.502	--	--	--
H ₂ O	34.645	34.043	2.313	1.748	1.921	0.101	68.613	65.676	55.527
K ₂ CO ₃	--	--	--	--	--	--	5.413	1.911	5.413
KHCO ₃	--	--	--	--	--	--	1.193	8.197	1.193
As ₂ O ₃	--	--	--	--	--	--	0.155	0.155	0.155
Total kg-mole/sec	57.446	57.446	25.716	21.602	21.408	3.650	75.377	75.989	62.291
Total kg/sec	888.48	888.48	317.37	153.13	153.13	155.89	2132.40	2296.64	1896.86

5-40

TABLE 5-15 (continued)

Stream No.	10	11	12	13	14	15	16	17	18
Component Description	Recycle Condensate	Recycle Condensate	Recycle Condensate	Makeup Water	Reformer Feed-water	Hydrogen Compressor Feed	Hydrogen Recycle	Hydrogen Product	Recycle Methane
Composition (kg-mole/sec)									
H ₂	--	--	--	--	--	13.828	5.998	7.609	0.221
CH ₄	--	--	--	--	--	5.659	0.287	0.363	5.009
CO	--	--	--	--	--	--	--	--	--
CO ₂	--	--	--	--	--	--	--	--	--
H ₂ O	0.464	31.730	1.921	7.105	41.220	0.124	--	--	--
K ₂ CO ₃	--	--	--	--	--	--	--	--	--
KHCO ₃	--	--	--	--	--	--	--	--	--
As ₂ O ₃	--	--	--	--	--	--	--	--	--
Total kg-mole/sec	0.464	31.730	1.921	7.105	41.220	19.611	6.285	7.972	5.230
Total kg/sec	8.35	571.11	34.58	127.88	741.92	120.78	16.68	21.14	80.73

5-40

The process steps through methanation are identical to the base case. The methanator effluent gas is cooled and compressed to 5.02 MN/m^2 (728 psia) by the hydrogen compressor (R-701) and fed to the cryogenic separation system for removal of methane. The methane-rich stream leaving the cryogenic separation system is recycled as reformer feed. The hydrogen product stream is compressed to a product delivery pressure of 10.34 MN/m^2 (1500 psia) by the hydrogen booster compressor (R-702). A portion of the hydrogen stream is compressed to 13.78 MN/m^2 (2000 psia) for recycle to the coal solution and hydrogasification reactors. Product hydrogen has a purity of 95.4%.

5.2.3. Case III, Process Temperature = 1144°K (1600°F)

A summary of the plant performance is given in Table 5-16, and details of power requirements and cooling water requirements are given in Tables 5-17 and 5-18. Overall stream flows are shown in Figs. 5-9 and 5-10 and Tables 5-19 and 5-20.

The principal modification to the base case [process temperature = 1033°K (1400°F)] was the use of two-stage shift converters with intermediate cooling. Less favorable equilibrium conditions for the water gas shift reaction in the reformer resulted in over twice the carbon monoxide in the reformer effluent compared with the base case. The temperature rise for single-stage shift conversion would have been too large. Shift conversion was accomplished in two stages, i.e., operating at high and low temperatures.

5.2.3.1. Hydrogen Production and Power Generation (Section 600). The detailed changes in process conditions from the base case are given in the process flow sheet shown in Fig. 5-11 and in the mass balance given in Table 5-21.

TABLE 5-16
OVERALL PERFORMANCE SUMMARY, CASE III, PROCESS TEMPERATURE = 1144°K (1600°F)

	Quantity at 100% Capacity	Gross Heat Value		% of Total Input
		(MW)	(10 ⁹ Btu/hr)	
Input				
Coal feed	195.66 kg/sec (18,640 tpd)	5682	19.40	65.5
Makeup water	238.92 kg/sec (3788 gpm)			
HTGR		3000	10.24	34.5
Total in		8682	29.64	100.0
Product				
Hydrogen	14.215 kg-mole/sec (972 MMscfd) 90.8 mol % ^(a)	4854	16.57	55.9
Aromatics	19.75 kg/sec (10,750 bpd)	838	2.86	9.6
Power	31 MW	31	0.11	0.4
Subtotal		5723	19.54	65.9
Waste Heat				
Char	35.8 kg/sec (3410 tpd)	668	2.28	7.7
To cooling water		2089	7.13	24.0
Other		202	0.69	2.3
Subtotal		2959	10.10	
Total Out		8682	29.64	100.0

(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

TABLE 5-17
POWER REQUIREMENTS, CASE III,
PROCESS TEMPERATURE = 1144°K (1600°F)

	Power (MW)
Section 600	
Power produced	
Circulator turbine	121
High-pressure turbine	118
Low-pressure turbine	203
Subtotal	442
Power consumed	
Helium circulator	(121)
Pumps	(15)
Subtotal	(136)
Net power produced, Section 600	306
Section 700	
Power produced	--
Power consumed	
Hydrogen compression	(237)
Pumps	(13)
Other	--
Subtotal	(250)
Net power consumed, Section 700	(250)
Net power consumed (Sections 100 to 500)	(25)
Total net power produced (Sections 100 to 700)	31

TABLE 5-18
COOLING WATER REQUIREMENTS, CASE III,
PROCESS TEMPERATURE = 1144°K (1600°F)

Item No.	Description	Heat Duty	
		MW	10 ⁹ Btu/hr
Sections 600, 700			
T-604	Condenser	196	0.670
T-706	Condenser	103	0.351
T-707	CO ₂ stripper condenser	888	3.031
T-708	HP condenser	71	0.244
R-701	Compressor intercoolers	147	0.501
Total Sections 600 to 700		1405 MW	(4797 MMBtu/hr)
Total Sections 100 to 500		684 MW	(2334 MMBtu/hr)
Total heat duty		2089 MW	(7131 MMBtu/hr)

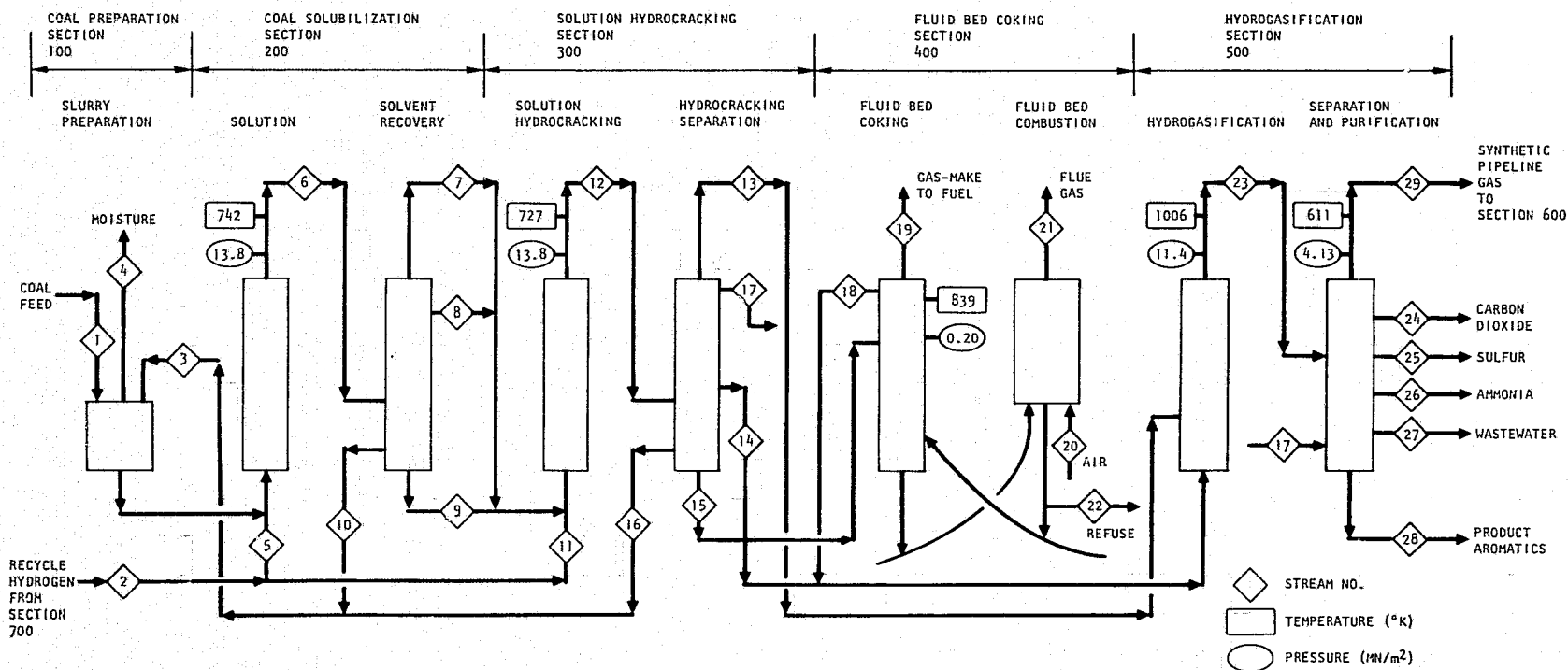


Fig. 5-9. Flow diagram, Sections 100-500, Case III

TABLE 5-19
HYDROGASIFICATION OF COAL LIQUIDS OVERALL PLANT FLOWS, CASE III, PROCESS TEMPERATURE = 1144°K (1600°F)

Stream No.	1	2	3	4	5	6	7	8	9	10
Component Description	Coal Feed	Recycle Hydrogen	Recycle Solvent	Moisture	Hydrogen to Coal Solution	Coal Solution Effluent	Hydrocarbon Gases	Distillate	Coal Liquids	Recycle Solvent
Composition (kg/sec)										
H ₂	--	23.95	--	--	11.99	9.80	9.80	--	--	--
C ₁ -C ₄	--	19.17	--	--	9.60	20.30	20.30	--	--	--
C ₅ -465°K	--	--	--	--	--	10.22	--	10.22	--	--
465°K-650°K	--	--	--	--	--	509.37	--	--	419.66	--
650°K-810°K	--	--	363.02	--	--		--	--		89.71
>810°K	--	--	--	--	--		--	--		--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	195.66	--	--	--	--	--	--	--	--	--
H ₂ O	--	0.22	--	10.14	0.11	20.55	20.55	--	--	--
H ₂ S	--	--	--	--	--			--	--	--
NH ₃	--	--	--	--	--			--	--	--
CO ₂	--	--	--	--	--			--	--	--
Air	--	--	--	--	--	--	--	--	--	--
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	195.66	43.34	363.02	10.14	21.70	570.24	50.65	10.22	419.66	89.71

TABLE 5-19 (continued)

Stream No.	11	12	13	14	15	16	17	18	19	20
Component Description	Hydrogen to Solution Hydro-cracking	Solution Hydro-cracker Effluent	Hydro-cracking Product Gas	Distillate	Heavy Coal Liquids	Recycle Solvent	Acid Gases	Coal Liquids	Gas-Make to Fuel	Combustion Air
Composition (kg/sec)										
H ₂	11.96	12.77	12.77	--	--	--	--	--	} 0.71	--
C ₁ -C ₄	9.57	50.41	50.41	--	--	--	--	--		--
C ₅ -465°K	--	} 81.22	--	} 81.22	--	--	--	--		--
465°K-650°K	--		--		--	--	--	--	--	--
650°K-810°K	--	} 331.29	--	--	} 57.98	273.31	--	19.19	--	--
>810°K	--		--	--		--	--	--	--	--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	--	--	--	--	--	--	--	--	--	--
H ₂ O	0.11	} 26.48	} 8.44	--	--	--	} 18.04	--	--	--
H ₂ S	--			--	--	--		--	--	--
NH ₃	--			--	--	--		--	--	--
CO ₂	--			--	--	--		--	--	--
Air	--	--	--	--	--	--	--	--	--	20.61
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	21.64	502.17	71.62	81.22	57.98	273.31	18.04	19.19	0.71	20.61

TABLE 5-19 (continued)

Stream No.	21	22	23	24	25	26	27	28	29	29	
Component Description	Flue Gas	Refuse	Hydro-gasifier Effluent	Carbon Dioxide	Sulfur	Ammonia	Waste-water	Product Aromatics	Reformer Feed Gas	Reformer Feed Composition (kg-mole/sec)	
Composition (kg/sec)											
H ₂	--	--	2.31	--	--	--	--	--	2.31	H ₂	1.145
C ₁ -C ₄	--	--	138.74	--	--	--	--	--	138.74	CH ₄	5.590
C ₅ -465°K	--	--	19.75	--	--	--	--	19.75	--	C ₂ H ₆	1.180
465°K-550°K	--	--	--	--	--	--	--	--	--	C ₃ H ₈	0.154
550°K-810°K	--	--	--	--	--	--	--	--	--	C ₄ H ₁₀	0.119
>810°K	--	--	--	--	--	--	--	--	--	Total	8.188
Coke (inc ash)	--	35.75	--	--	--	--	--	--	--	(kg-mole/sec)	
Coal	--	--	--	--	--	--	--	--	--		
H ₂ O	--	--	11.23	--	--	--	--	--	--		
H ₂ S	--	--		--	4.99	--	17.99	--	--		
NH ₃	--	--		--	--	3.03	--	--	--		
CO ₂	--	--		3.26	--	--	--	--	--		
Air	--	--	--	--	--	--	--	--	--		
Flue Gas	22.94	--	--	--	--	--	--	--	--		
Total kg/sec	22.94	35.75	172.03	3.26	4.99	3.03	17.99	19.75	141.05		

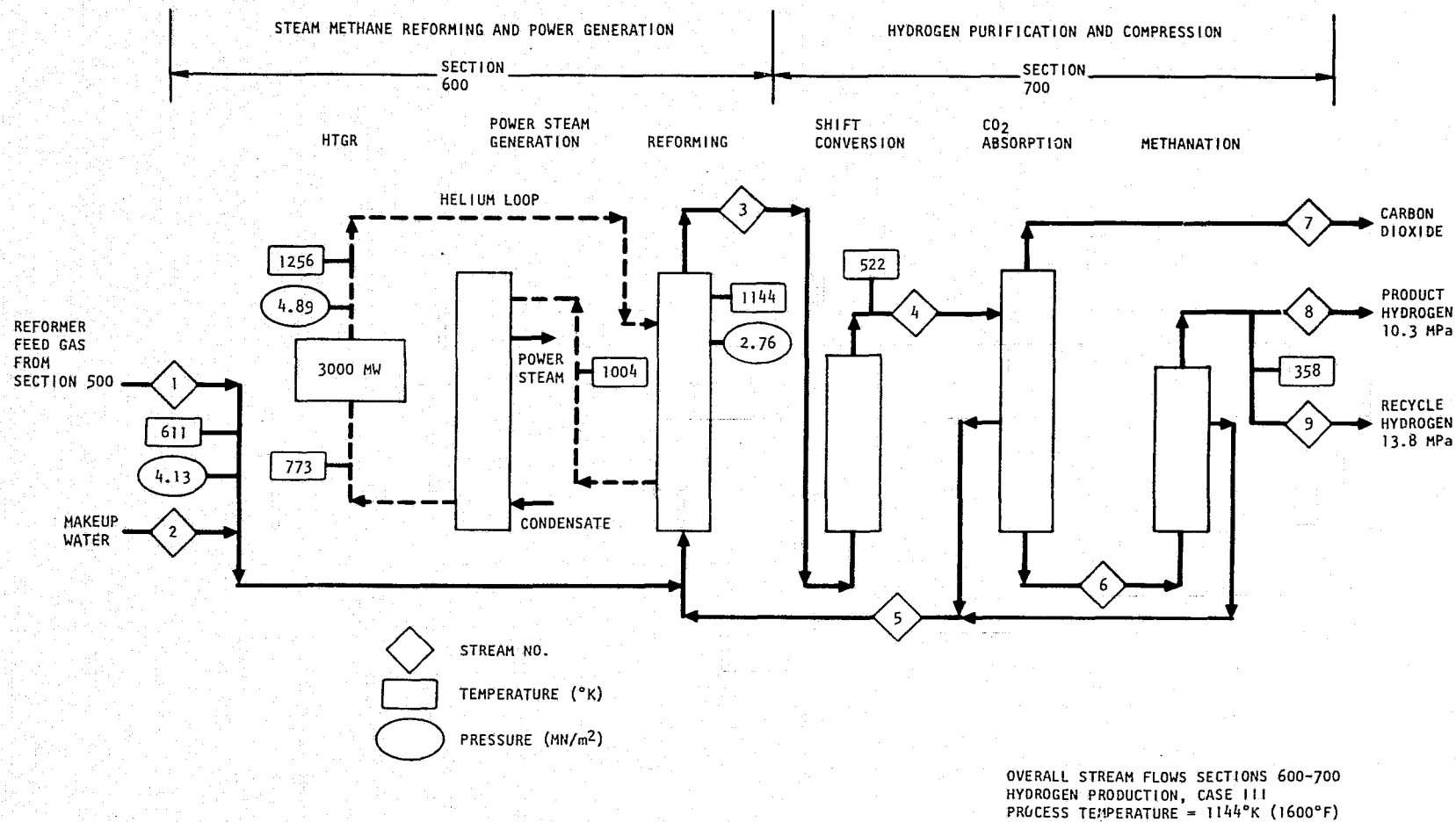
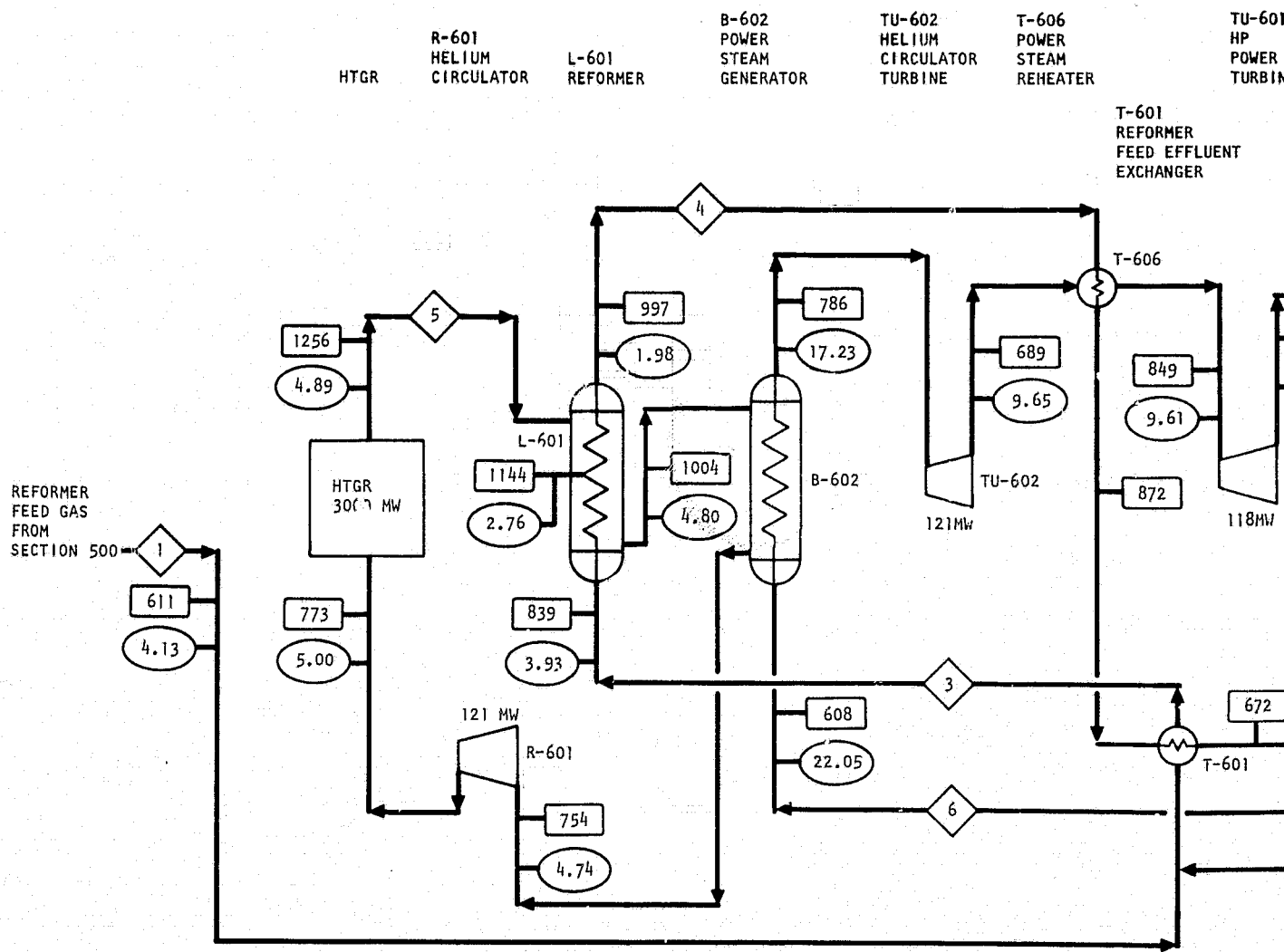


Fig. 5-10. Flow diagram, Sections 600-700, Case III

TABLE 5-20
HYDROGEN PRODUCTION AND POWER GENERATION OVERALL STREAM FLOWS, CASE III, PROCESS TEMPERATURE = 1144°K (1600°F)

Stream No.	1	2	3	4	5	6	7	8	9	Stream No.	8	9
Component Description	Reformer Feed Gas	Makeup Water	Reformer Effluent	Shift Converter Effluent	Recycle Condensate	Carbon Dioxide Absorber Effluent	Carbon Dioxide	Recycle Hydrogen to Coal Liquefaction	Product Hydrogen	Hydrogen Composition (kg/sec)	Recycle Hydrogen	Product Hydrogen
Composition (kg-mole/sec)												
H ₂	1.145	--	22.181	26.008	--	25.924	0.084	11.891	12.905	H ₂	23.95	25.99
CH ₄	5.590	--	2.150	2.150	--	2.150	--	1.196	1.297	CH ₄	19.17	20.79
C ₂ H ₆	1.180	--	--	--	--	--	--	--	--	CO	--	--
C ₃ H ₈	0.154	--	--	--	--	--	--	--	--	CO ₂	--	--
C ₄ H ₁₀	0.119	--	--	--	--	--	--	--	--	H ₂ O	0.22	0.23
CO	--	--	4.071	0.244	--	0.244	--	--	--	Total kg/sec	43.34	47.01
CO ₂	--	--	2.667	6.494	--	0.099	6.395	--	--			
H ₂ O	--	13.274	17.260	13.433	13.391	2.245	0.459	0.012	0.013			
Total kg-mole/sec	8.188	13.274	48.329	48.329	13.391	30.662	6.938	13.099	14.215			
kg/sec	141.05	238.92	620.99	620.99	241.02	138.26	289.62	43.34	47.01			



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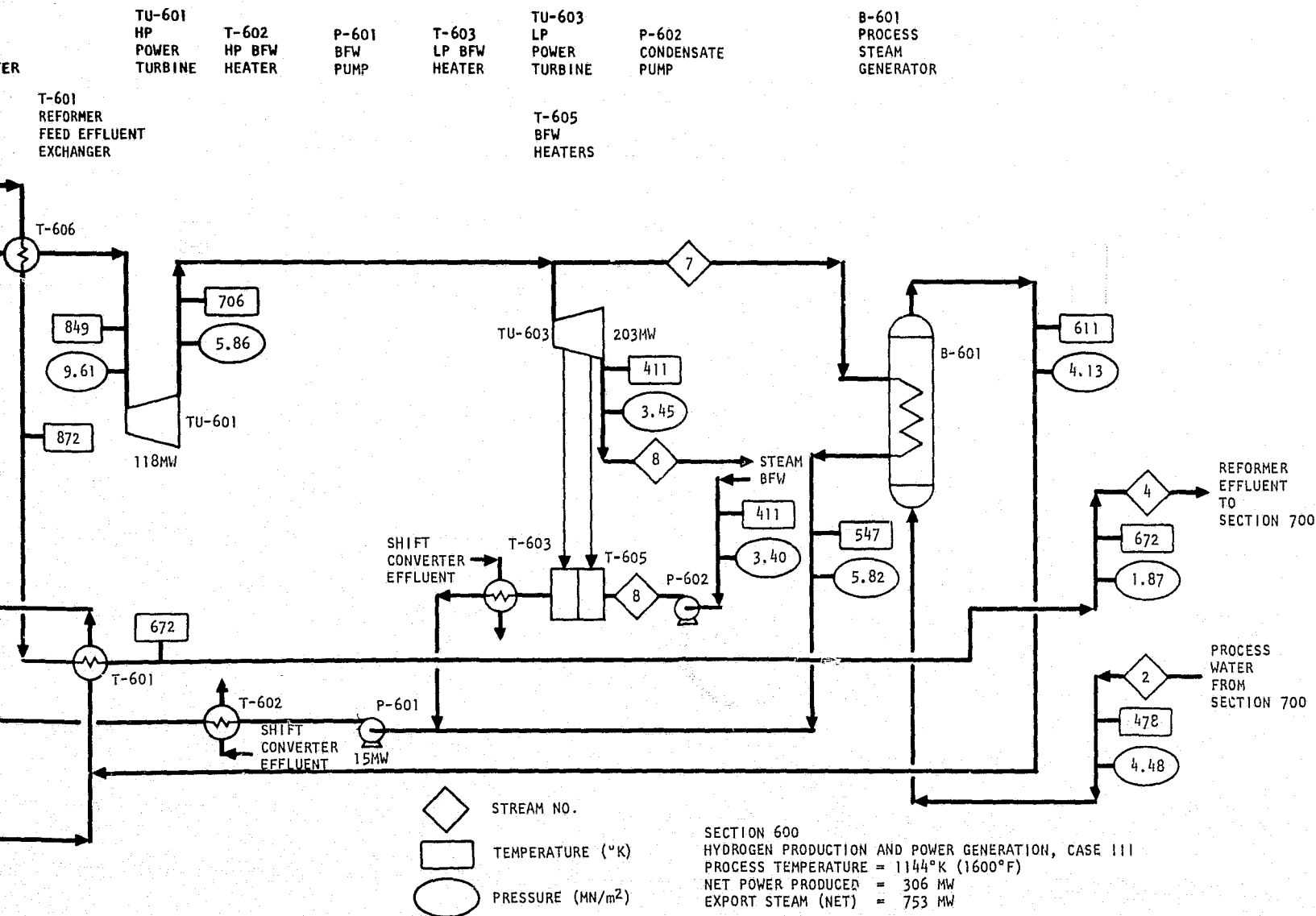


Fig. 5-11. Flow diagram, Section 600, Case III

TABLE 5-21

HYDROGEN PRODUCTION AND POWER GENERATION STREAM FLOWS, CASE III, PROCESS TEMPERATURE = 1144°K (1600°F)

Stream No.	1	2	3	4	5	6	7	8
Component Description	Reformer Feed Gas	Reformer Feed-water	Reformer Feed	Reformer Effluent	Helium Heat Transfer Loop	Boiler Feed-Water	Steam to Process Steam Generator	Steam From LP Turbine
Composition (kg-mole/sec)								
He	--	--	--	--	--	--	--	--
H ₂	1.145	--	1.145	22.181	--	--	--	--
CH ₄	5.590	--	5.590	2.150	--	--	--	--
C ₂ H ₆	1.180	--	1.180	--	--	--	--	--
C ₃ H ₈	0.154	--	0.154	--	--	--	--	--
C ₄ H ₁₀	0.119	--	0.119	--	--	--	--	--
CO	--	--	--	4.071	--	--	--	--
CO ₂	--	--	--	2.667	--	--	--	--
H ₂ O	--	26.665	26.665	17.260	298.792	49.423	28.492	19.447
Total kg-mole/sec	8.188	26.665	34.853	48.329	298.792	49.423	28.492	19.447
Total kg/sec	141.05	479.94	620.99	620.99	1197.05	889.56	512.82	350.03

Hot helium at 1256°K (1800°F) transfers heat from the HTGR nuclear core to the reformers and steam generators. The distribution of HTGR thermal power is as follows:

	<u>Thermal Power (MW)</u>
Reformer	1567
Steam generator	<u>1554</u>
Subtotal	3121
Helium circulators	<u>-121</u>
Total HTGR power	3000

Process steps through reforming and the reformer feed-effluent exchanger (T-601) are similar to the base case. Catalyst bed discharge conditions are 1144°K (1600°F) and 2.76 MN/m^2 (400 psia), and the process gas exits through integral return tubes leaving the reformer at 997°K (1335°F). After heat exchange with the reformer feed, the process gas at 672°K (750°F) is sent forward to Section 700 for purification.

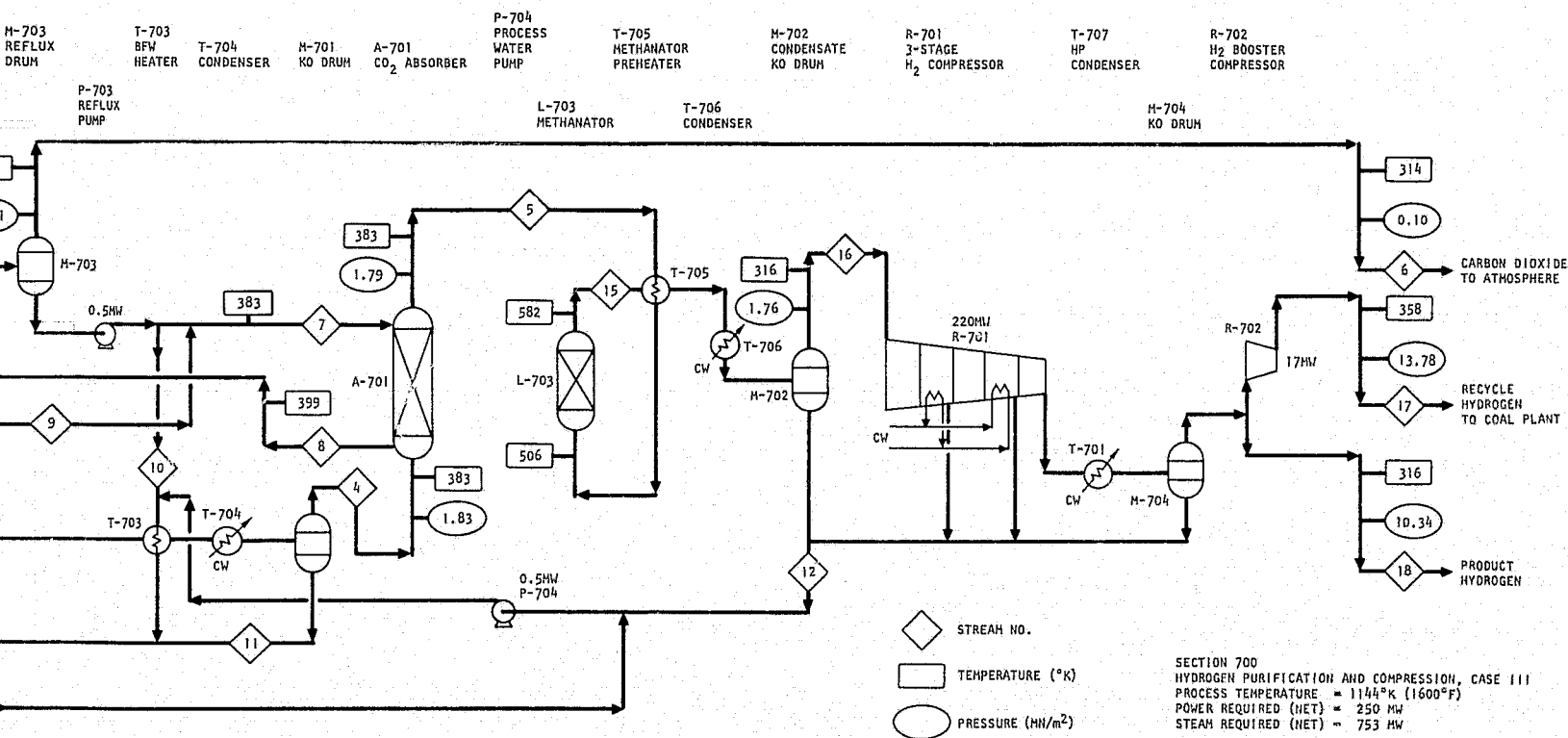
Minor changes were required in the power steam cycle. The low-pressure power turbine (TU-603) exhaust pressure was increased to 0.34 MN/m^2 (50 psia), and the exhaust steam was exported to Section 700 to supply heat requirements for the CO_2 stripper. Boiler feedwater preheat was supplied by the first-stage shift converter effluent (Section 700).

5.2.3.2. Hydrogen Purification and Compression (Section 700). The detailed changes in process conditions from the base case are given in the process flow sheet shown in Fig. 5-12 and the mass balance given in Table 5-22.

The reformer effluent at 672°K (750°F) is fed to the first-stage shift converters for conversion of a portion of the remaining carbon monoxide by the water gas shift reaction. The first-stage shift converter effluent is cooled by heat exchange with the power cycle boiler feedwater (exchangers T-602, T-603). Conversion of most of the remaining carbon monoxide is



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Fig. 5-12. Flow diagram, Section 700, Case III

TABLE 5-22
HYDROGEN PURIFICATION AND COMPRESSION STREAM FLOWS, CASE III, PROCESS TEMPERATURE = 1144°K (1600°F)

Stream No.	1	2	3	4	5	6	7	8	9
Component Description	Reformer Effluent	1st-Stage Shift Converter Effluent	2nd-Stage Shift Converter Effluent	Carbon Dioxide Absorber Feed	Carbon Dioxide Absorber Effluent	Carbon Dioxide to Atmosphere	Lean Solution to Absorber	Rich Solution to Stripper	Lean Solution From Stripper
Composition (kg-mole/sec)									
H ₂	22.181	24.217	26.008	26.008	25.924	0.084	0.004	0.088	0.004
CH ₄	2.150	2.150	2.150	2.150	2.150	--	--	--	--
CO	4.071	2.035	0.244	0.244	0.244	--	--	--	--
CO ₂	2.667	4.703	6.494	6.494	0.099	6.395	--	--	--
H ₂ O	17.260	15.224	13.433	3.012	2.245	0.459	120.979	115.351	103.589
K ₂ CO ₃	--	--	--	--	--	--	9.592	3.197	9.592
KHCO ₃	--	--	--	--	--	--	2.133	14.923	2.133
As ₂ O ₃	--	--	--	--	--	--	0.273	0.273	0.273
Total kg-mole/sec	48.329	48.329	48.329	37.908	30.662	6.938	132.981	133.832	115.591
Total kg/sec	620.99	620.99	620.99	433.42	138.26	289.62	3769.36	4064.52	3456.36

TABLE 5-22 (continued)

Stream No.	10	11	12	13	14	15	16	17	18
Component Description	Recycle Condensate	Recycle Condensate	Recycle Condensate	Makeup Water	Reformer Feed-water	Methanator Effluent	Compressor Feed	Hydrogen Recycle	Hydrogen Product
Composition (kg-mole/sec)									
H ₂	--	--	--	--	--	24.796	24.796	11.891	12.905
CH ₄	--	--	--	--	--	2.493	2.493	1.196	1.297
CO	--	--	--	--	--	--	--	--	--
CO ₂	--	--	--	--	--	--	--	--	--
H ₂ O	0.308	10.421	2.662	13.274	26.665	2.687	0.147	0.012	0.013
K ₂ CO ₃	--	--	--	--	--	--	--	--	--
KHCO ₃	--	--	--	--	--	--	--	--	--
As ₂ O ₃	--	--	--	--	--	--	--	--	--
Total kg-mole/sec	0.308	10.421	2.662	13.274	26.665	29.976	27.436	13.099	14.215
Total kg/sec	5.54	187.57	47.91	238.92	479.94	138.26	92.55	43.34	47.01

accomplished in the second-stage shift converters. The remaining process steps are similar to the base case.

Additional low-pressure steam from Section 600 was required to supply the heat requirements for the CO₂ stripper.

Product hydrogen has a purity of 90.8%.

5.2.4. Case IV, Process Temperature = 1256°K (1800°F)

A summary of the plant performance is given in Table 5-23, and additional details on power requirements and cooling water requirements are given in Tables 5-24 and 5-25. Overall stream flows are given in Figs. 5-13 and 5-14 and Tables 5-26 and 5-27.

The principal modification to the base case was the use of a Purisol CO₂ removal process in place of a hot carbonate process. The helium loop heat balance taken from the previous high-temperature heat source study (Ref. 5-1) did not leave sufficient power for steam generation to meet the process heat requirements if CO₂ removal by the hot carbonate process was used. The Purisol process (licensed by Ralph M. Parsons Co. in the U.S.) offers a substantial reduction in process heat requirements over the more conventional hot carbonate process.

In addition, shift conversion was accomplished in two stages with intermediate cooling. The relatively large quantities of carbon monoxide in the reformer effluent would give too large a temperature rise in single-stage shift converters.

5.2.4.1. Hydrogen Production and Compression (Section 600). The detailed changes in process conditions from the base case are given in the flow sheet shown in Fig. 5-15 and the mass balance given in Table 5-28.

TABLE 5-23
OVERALL PERFORMANCE SUMMARY, CASE IV, PROCESS TEMPERATURE = 1256°K (1800°F)

	Quantity at 100% Capacity	Gross Heat Value		% of Total Input
		(MW)	(10 ⁹ Btu/hr)	
Input				
Coal feed	210.75 kg/sec (20,080 tpd)	6107	20.85	67.1
Makeup water	262.23 kg/sec (4158 gpm)			
HTGR	3000 MW	3000	10.24	32.9
Total in		9107	31.09	100.0
Product				
Hydrogen	15.244 kg-mole/sec (1042 MMscfd) 91.0 mol % ^(a)	5173	17.66	56.8
Aromatics	21.38 kg/sec (11,640 bpd)	905	3.09	9.9
Power	18 MW	18	0.06	0.2
Subtotal		6096	20.81	66.9
Waste heat				
Char	38.49 kg/sec (3666 tpd)	718	2.45	7.9
To cooling water		2056	7.02	22.6
Other		237	0.81	2.6
Subtotal		3011	10.28	33.1
Total out		9107	31.09	100.0

(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

TABLE 5-24
POWER REQUIREMENTS, CASE IV,
PROCESS TEMPERATURE = 1256°K (1800°F)

	Power (MW)
Section 600	
Power produced	
Circulator turbine	102
High-pressure turbine	74
Low-pressure turbine	274
Subtotal	450
Power consumed	
Helium circulator	(102)
Pumps	(14)
Subtotal	(116)
Net power produced, Section 600	334
Section 700	
Power produced	32
Power consumed	
Hydrogen compression	(246)
Pumps	(75)
Other	--
Subtotal	(321)
Net power consumed, Section 700	(289)
Net power consumed (Sections 100 to 500)	(27)
Total net power produced (Sections 100 to 700)	18

TABLE 5-25
COOLING WATER REQUIREMENTS, CASE IV,
PROCESS TEMPERATURE = 1256°K (1800°F)

Item No.	Description	Heat Duty	
		MW	10 ⁹ Btu/hr
Sections 600, 700			
T-604	Power steam condenser	518	1.770
T-203	Absorber precooler	114	0.389
T-705	Compressor precooler	117	0.398
T-706	Compressor precooler	460	1.570
R-701	Compressor intercoolers	114	0.389
Total Sections 600 to 700		1323 MW	(4516 MMBtu/hr)
Total Sections 100 to 500		733 MW	(2504 MMBtu/hr)
Total heat duty		2056 MW	(7020 MMBtu/hr)

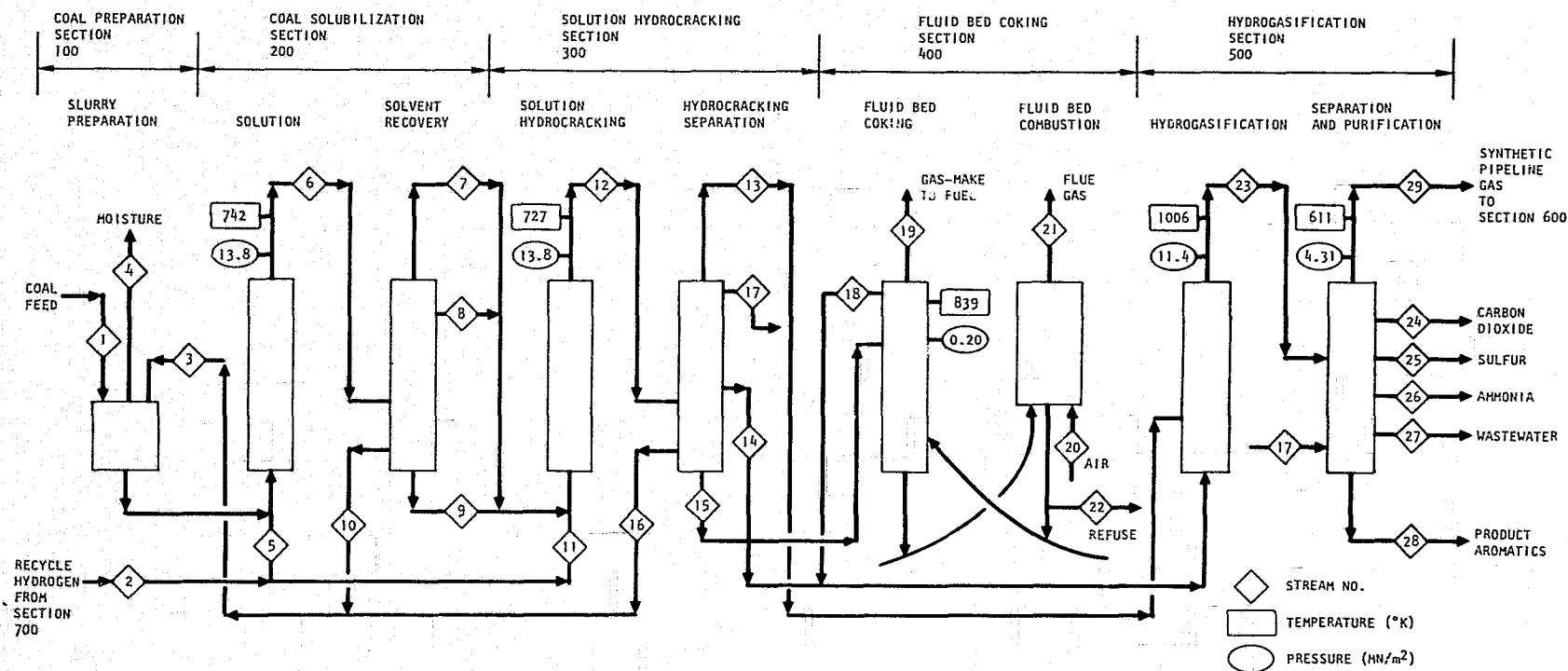


Fig. 5-13. Flow diagram, Sections 100-500, Case IV

TABLE 5-26
HYDROGASIFICATION OF COAL LIQUIDS OVERALL PLANT FLOWS, CASE IV, PROCESS TEMPERATURE = 1256°K (1800°F)

Stream No.	1	2	3	4	5	6	7	8	9	10
Component Description	Coal Feed	Recycle Hydrogen	Recycle Solvent	Moisture	Hydrogen to Coal Solution	Coal Solution Effluent	Hydrocarbon Gases	Distillate	Coal Liquids	Recycle Solvent
Composition (kg/sec)										
H ₂	--	25.79	--	--	12.91	10.59	10.59	--	--	--
C ₁ -C ₄	--	20.00	--	--	10.01	21.50	21.50	--	--	--
C ₅ -465°K	--	--	--	--	--	11.01	--	11.01	--	--
465°K-650°K	--	--	--	--	--	548.65	--	--	452.03	--
650°K-810°K	--	--	391.02	--	--		--	--		96.62
>810°K	--	--	--	--	--		--	--		--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	210.75	--	--	--	--	--	--	--	--	--
H ₂ O	--	0.23	--	10.93	0.12	22.13	22.13	--	--	--
H ₂ S	--	--	--	--	--			--	--	--
NH ₃	--	--	--	--	--			--	--	--
CO ₂	--	--	--	--	--	--	--	--	--	--
Air	--	--	--	--	--	--	--	--	--	--
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	210.75	46.02	391.02	10.93	23.04	613.88	54.22	11.01	452.03	96.62

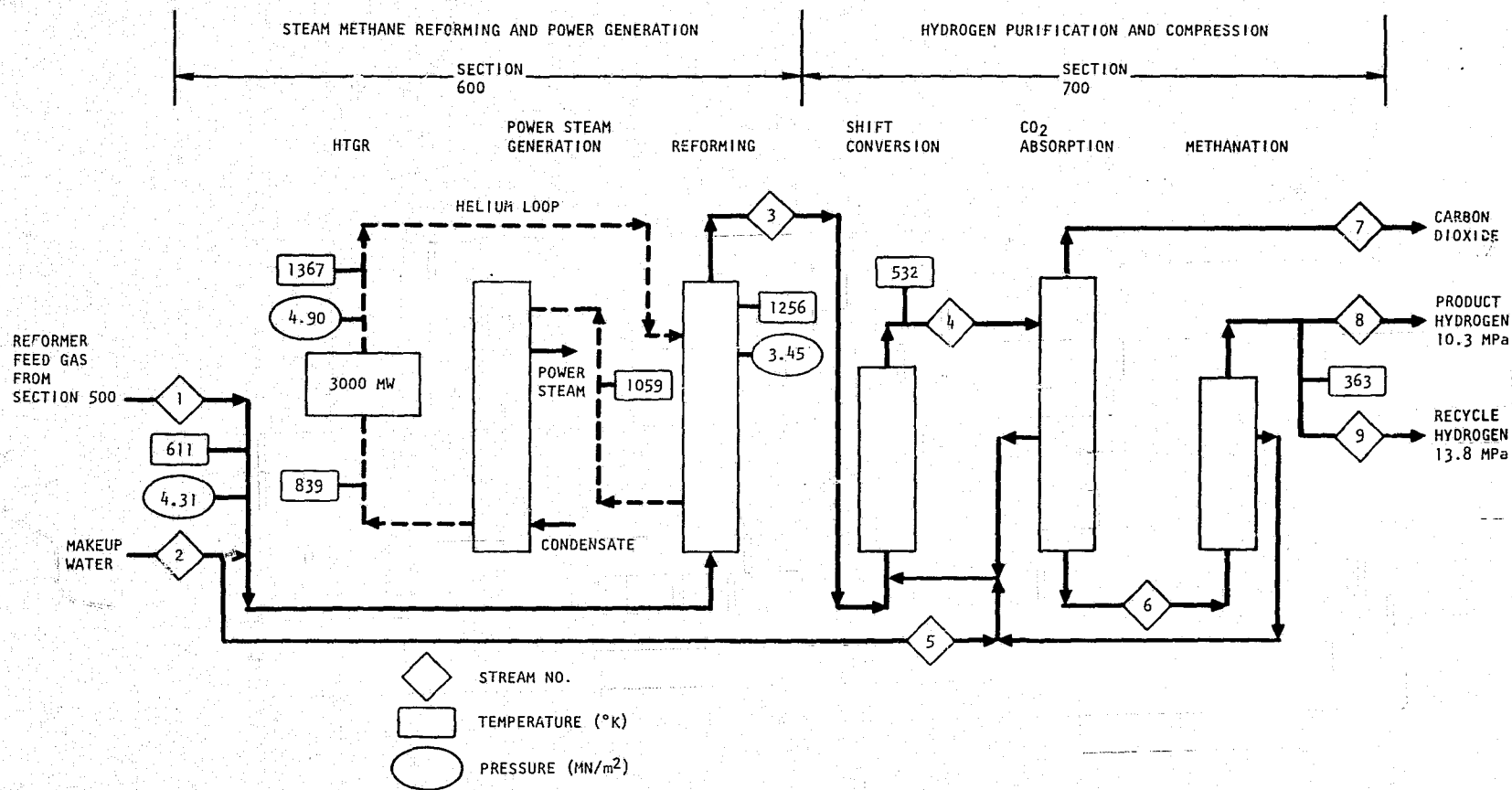
5-64

TABLE 5-26 (continued)

Stream No.	11	12	13	14	15	16	17	18	19	20
Component Description	Hydrogen to Solution Hydro-cracking	Solution Hydro-cracker Effluent	Hydro-cracking Product Gas	Distillate	Heavy Coal Liquids	Recycle Solvent	Acid Gases	Coal Liquids	Gas-Make to Fuel	Combustion Air
Composition (kg/sec)										
H ₂	12.88	13.72	13.72	--	--	--	--	--	} 0.77	--
C ₁ -C ₄	9.99	53.68	53.68	--	--	--	--	--		--
C ₅ -465°K	--	} 87.48	--	} 87.48	--	--	--	--		--
465°K-650°K	--		--		--	--	--	--	--	--
650°K-810°K	--	} 356.84	--	--	} 62.44	294.40	--	20.67	--	--
>810°K	--		--	--		--	--	--	--	--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	--	--	--	--	--	--	--	--	--	--
H ₂ O	0.11	} 28.52	} 9.10	--	--	--	} 19.42	--	--	--
H ₂ S	--			--	--	--		--	--	--
NH ₃	--			--	--	--		--	--	--
CO ₂	--			--	--	--		--	--	--
Air	--	--	--	--	--	--	--	--	--	22.20
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	22.98	540.24	76.50	87.48	62.44	294.40	19.42	20.67	0.77	22.20

TABLE 5-26 (continued)

Stream No.	21	22	23	24	25	26	27	28	29	29	
Component Description	Flue Gas	Refuse	Hydro-gasifier Effluent	Carbon Dioxide	Sulfur	Ammonia	Waste-water	Product Aromatics	Reformer Feed Gas	Reformer Feed Composition (kg-mole/sec)	
Composition (kg/sec)											
H ₂	--	--	2.48	--	--	--	--	--	2.48	H ₂	1.233
C ₁ -C ₄	--	--	148.70	--	--	--	--	--	148.70	CH ₄	5.972
C ₅ -465°K	--	--	21.38	--	--	--	--	21.38	--	C ₂ H ₆	1.271
465°K-650°K	--	--	--	--	--	--	--	--	--	C ₃ H ₈	0.166
650°K-810°K	--	--	--	--	--	--	--	--	--	C ₄ H ₁₀	0.129
>810°K	--	--	--	--	--	--	--	--	--		
Coke (inc ash)	--	38.49	--	--	--	--	--	--	--	Total	8.771
Coal	--	--	--	--	--	--	--	--	--	(kg-mole/sec)	
H ₂ O	--	--	12.09	--	--	--	19.37	--	--		
H ₂ S	--	--		--	5.37	--	--	--	--		
NH ₃	--	--		--	--	3.26	--	--	--		
CO ₂	--	--		3.51	--	--	--	--	--		
Air	--	--	--	--	--	--	--	--	--		
Flue Gas	24.71	--	--	--	--	--	--	--	--		
Total kg/sec	24.71	38.49	184.65	3.51	5.37	3.26	19.37	21.38	151.18		



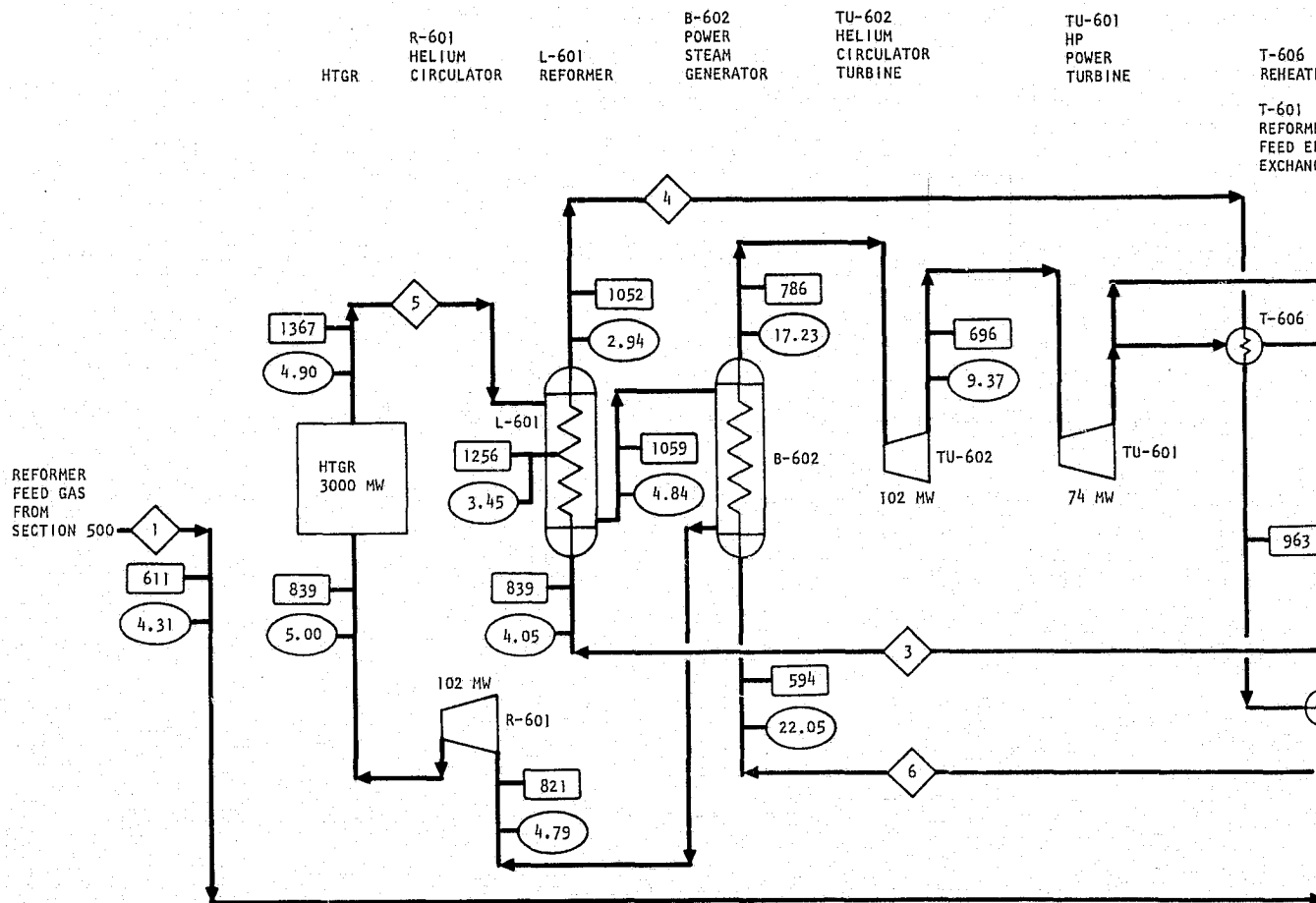
OVERALL STREAM FLOWS SECTIONS 600-700
HYDROGEN PRODUCTION, CASE IV
PROCESS TEMPERATURE = 1256°K (1800°F)

Fig. 5-14. Flow diagram, Sections 600-700, Case IV

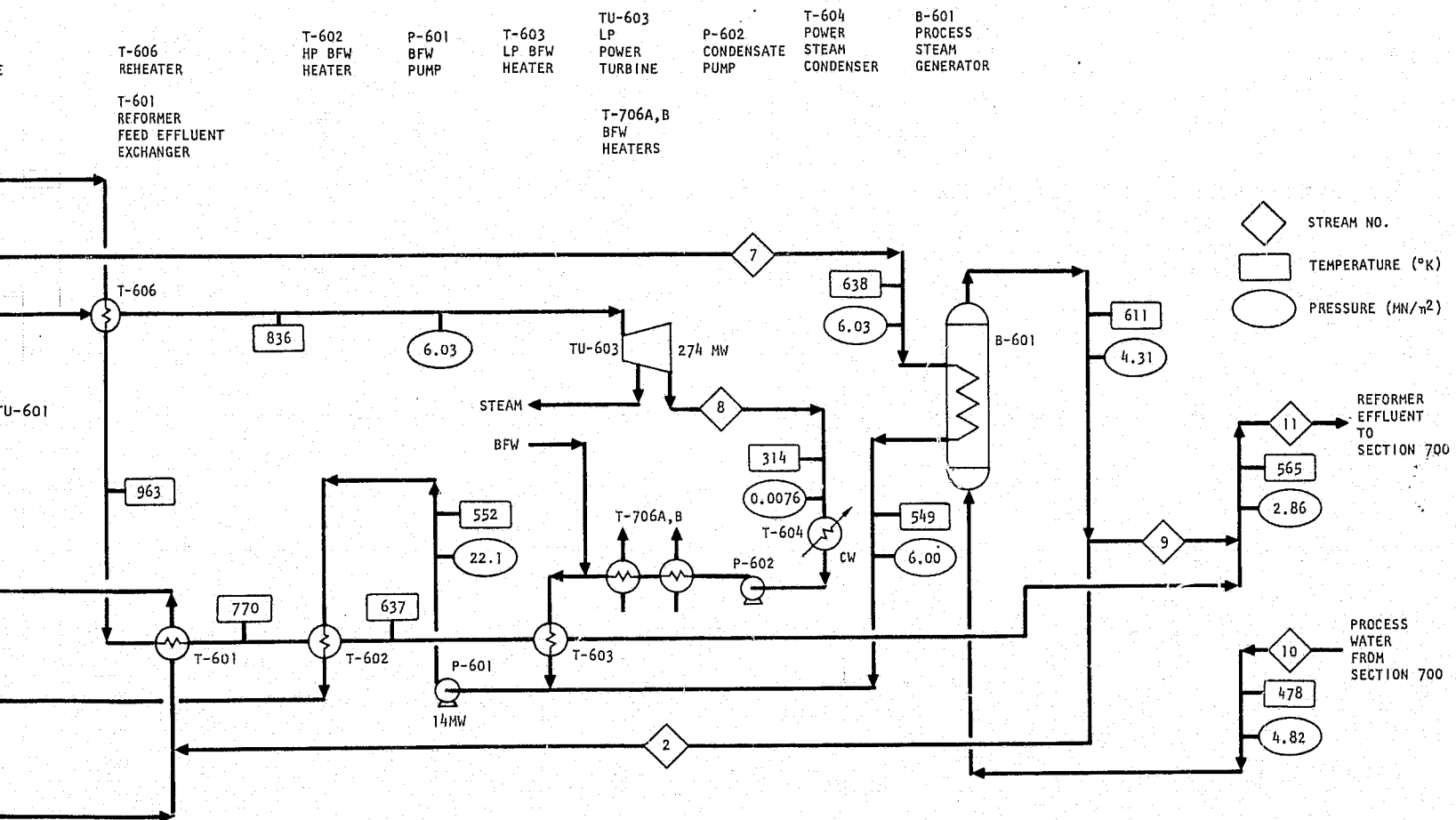
TABLE 5-27

HYDROGEN PRODUCTION AND POWER GENERATION OVERALL STREAM FLOWS, CASE IV, PROCESS TEMPERATURE = 1256°K (1800°F)

Stream No.	1	2	3	4	5	6	7	8	9	Stream No.	8	9
Component Description	Reformer Feed Gas	Makeup Water	Reformer Effluent	Shift Converter Effluent	Recycle Condensate	Carbon Dioxide Absorber Effluent	Carbon Dioxide	Recycle Hydrogen to Coal Liquefaction	Product Hydrogen	Hydrogen Composition (kg/sec)	Recycle Hydrogen	Product Hydrogen
Composition (kg-mole/sec)												
H ₂	1.233	--	22.267	28.294	--	28.035	0.259	12.807	13.877	H ₂	25.79	27.95
CH ₄	5.972	--	2.192	2.192	--	2.192	--	1.248	1.353	CH ₄	20.00	21.69
C ₂ H ₆	1.271	--	--	--	--	--	--	--	--	CO	--	--
C ₃ H ₈	0.166	--	--	--	--	--	--	--	--	CO ₂	--	--
C ₄ H ₁₀	0.129	--	--	--	--	--	--	--	--	H ₂ O	0.23	0.25
CO	--	--	6.320	0.293	--	0.285	0.008	--	--	Total kg/sec	46.02	49.89
CO ₂	--	--	1.016	7.043	--	0.124	7.549	--	--			
H ₂ O	--	14.569	5.940	16.599	0.277	0.026	0.696	0.013	0.014			
Total kg-mole/sec	8.771	14.569	37.735	54.421	0.277	30.662	8.512	14.068	15.244			
kg/sec	151.18	262.23	408.42	708.76	4.99	105.49	317.50	46.02	49.89			



EOLDOUT FRAME



SECTION 600
 HYDROGEN PRODUCTION AND POWER GENERATION, CASE IV
 PROCESS TEMPERATURE = 1256°K
 NET POWER PRODUCED = 334 MW
 EXPORT STEAM (NET) = 18 MW

FOLDOUT FRAME 2

Fig. 5-15. Flow diagram, Section 600, Case IV

TABLE 5-28
HYDROGEN PRODUCTION AND POWER GENERATION STREAM FLOWS, CASE IV, PROCESS TEMPERATURE = 1256°K (1800°F)

Stream No.	1	2	3	4	5	6	7	8	9	10	11
Component Description	Reformer Feed Gas	Reformer Steam	Reformer Feed	Reformer Effluent	Helium Loop	Power Cycle BFW	Steam to Process Steam Generator	Steam From LP Turbine	Steam to 1st-Stage Shift Converter	Process BFW	1st-Stage Shift Converter Feed
Composition (kg-mole/sec)											
He	--	--	--	--	273.418	--	--	--	--	--	--
H ₂	1.233	--	1.233	22.267	--	--	--	--	--	--	22.267
CH ₄	5.972	--	5.972	2.192	--	--	--	--	--	--	2.192
C ₂ H ₆	1.271	--	1.271	--	--	--	--	--	--	--	--
C ₃ H ₈	0.166	--	0.166	--	--	--	--	--	--	--	--
C ₄ H ₁₀	0.129	--	0.129	--	--	--	--	--	--	--	--
CO	--	--	--	6.320	--	--	--	--	--	--	6.320
CO ₂	--	--	--	1.016	--	--	--	--	--	--	1.016
H ₂ O	--	14.292	14.292	5.940	--	40.392	27.091	12.944	8.833	23.125	14.773
Total kg-mole/sec	8.771	14.292	23.063	37.735	273.418	40.392	27.091	12.944	8.833	23.125	46.568
Total kg/sec	151.18	257.24	408.42	408.42	1093.38	727.02	487.61	232.98	158.99	416.23	567.41

Hot helium at 1367°K (2000°F) carries heat from the HTGR nuclear core to the reformers and steam generators. The distribution of HTGR thermal power is as follows:

	<u>Thermal Power (MW)</u>
Reformers	1746
Steam generators	<u>1356</u>
Subtotal	3102
Helium circulators	<u>-102</u>
Total HTGR power	3000

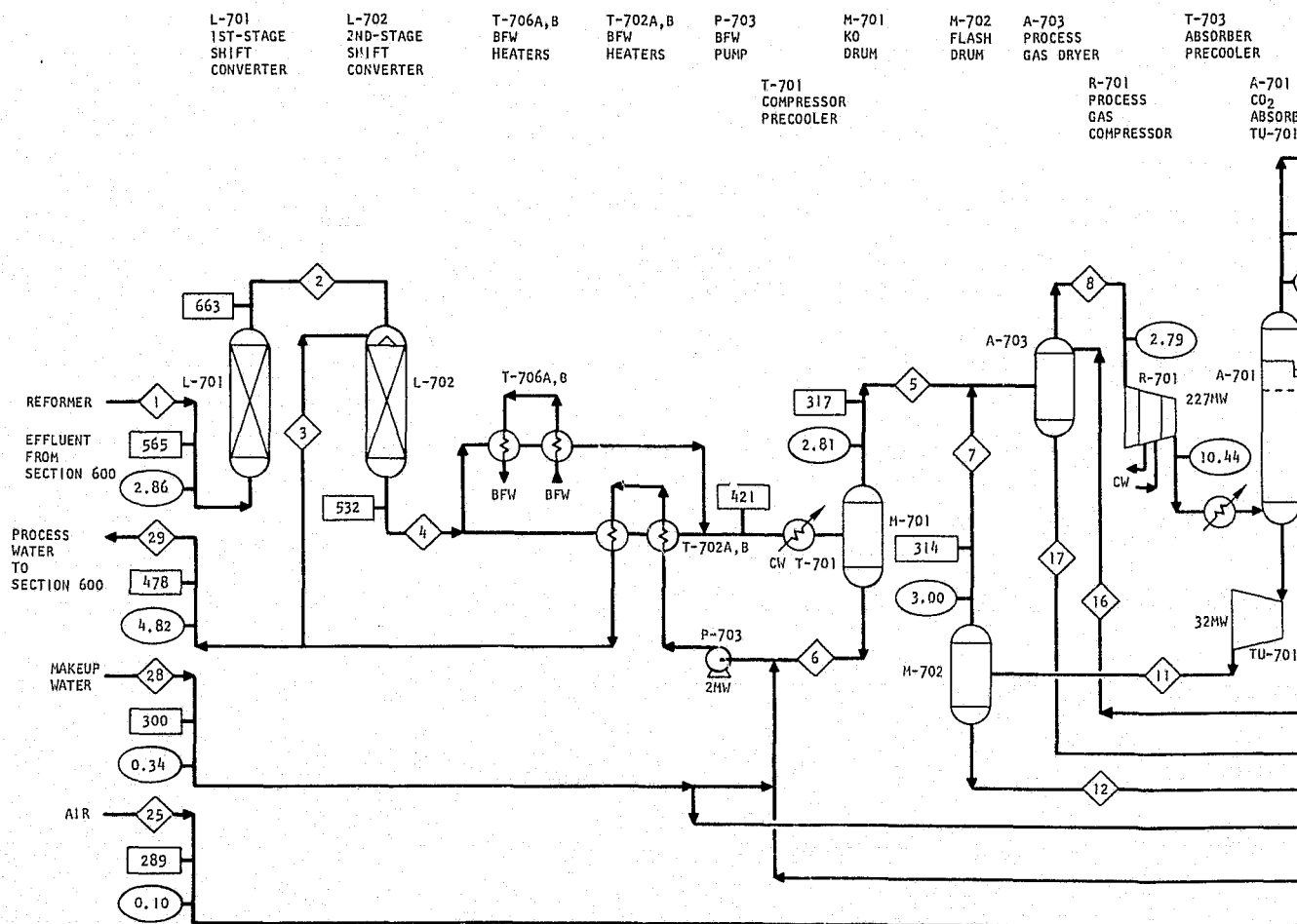
The process steps for reforming process gas and process gas heat recovery are identical to the base case. Catalyst bed discharge conditions are 1256°K (1800°F) and 3.45 MN/m² (500 psia), and the process gas exits through integral return tubes leaving the reformer at 1052°K (1433°F).

Minor changes in the power steam cycle were required. The exhaust steam from the helium circulator turbine (TU-602) was directly fed to the high-pressure power turbine (TU-601) without reheat. The majority of the high-pressure power turbine exhaust was condensed in the process steam generator (B-601). The remaining steam was reheated and fed to the low-pressure power turbine. Extraction steam from the low-pressure power turbine was exported to Section 700 to supply the heat requirement for solvent drying in the CO₂ removal system.

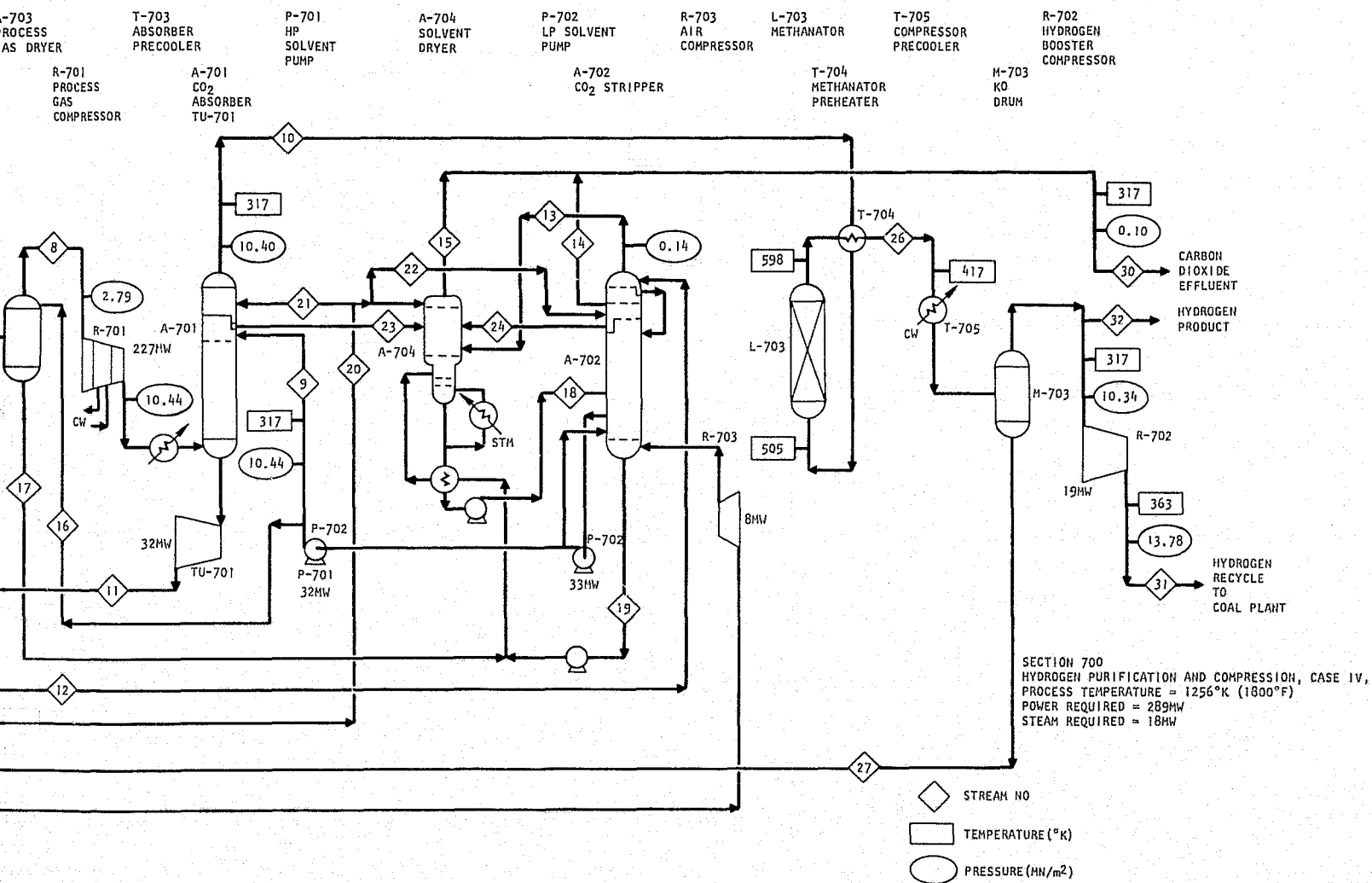
5.2.4.2. Hydrogen Purification and Compression (Section 700). The following process description refers to the process flow sheet shown in Fig. 5-16 and the mass balance given in Table 5-29.

Reformer effluent at 558°K (544°F) feeds the shift converters (L-701, L-702). Shift conversion is accomplished in two stages with an intermediate quench with recycled hot condensate. Excess steam in the shift converter effluent is condensed in exchangers T-702 and T-706 and condenser T-701. Condensate is removed in the knockout drum (M-701) and recycled for treatment. The process gas is fed to the CO₂ removal system.

REPRODUCIBILITY OF THE
ORIGINAL PAGE IS POOR



FOLDOUT FRAME



FOLDOUT FRAME 2

Fig. 5-16. Flow diagram, Section 700, Case IV

TABLE 5-29

HYDROGEN PURIFICATION AND COMPRESSION STREAM FLOWS, CASE IV, PROCESS TEMPERATURE = 1256°K (1800°F)

Stream No.	1	2	3	4	5	6	7	8
Component Description	1st-Stage Shift Converter Feed	1st-Stage Shift Converter Effluent	Condensate to 2nd-Stage Shift Converter	2nd-Stage Shift Converter Effluent	Feed Gas to CO ₂ Removal System	Recycle Condensate	Recycle Hydrogen	CO ₂ Absorber Feed
Composition (kg-mole/sec)								
H ₂	22.267	26.337	--	28.294	28.294	--	0.612	28.904
CH ₄	2.192	2.192	--	2.192	2.192	--	--	2.192
CO	6.320	2.250	--	0.293	0.293	--	0.010	0.303
CO ₂	1.016	5.086	--	7.043	7.043	--	0.511	7.504
Air	--	--	--	--	--	--	--	--
H ₂ O	14.773	10.703	7.853	16.599	0.120	16.479	--	0.009
NMP	--	--	--	--	--	--	--	--
Total kg-mole/sec	46.568	46.568	7.853	54.421	37.942	16.479	1.133	38.912
Total kg/sec	567.41	567.41	141.35	708.76	412.15	296.61	23.98	431.93

TABLE 5-29 (continued)

Stream No.	9	10	11	12	13	14	15	16	17
Component Description	Lean Solution to Absorber	Absorber Effluent Gas	Rich Solution to Flash Drum	Solvent to CO ₂ Stripper	LP Flash Gas	CO ₂ Stripper Effluent Gas	Solvent Dryer Effluent Gas	Solvent to Dryer	Solvent Effluent From Dryer
Composition (kg-mole/sec)									
H ₂	--	28.035	0.869	0.257	0.257	--	0.259	--	0.002
CH ₄	--	2.192	--	--	--	--	--	--	--
CO	--	0.285	0.018	0.008	0.008	--	0.008	--	--
CO ₂	0.050	0.124	7.430	6.919	5.134	1.735	5.184	--	0.050
Air	--	--	--	--	--	2.290	--	--	--
H ₂ O	1.629	0.026	1.638	1.638	0.008	0.262	0.502	0.013	0.124
NMP	51.131	--	51.131	51.131	--	--	--	0.393	0.393
Total kg-mole/sec	52.810	30.662	61.086	59.953	5.407	4.287	5.953	0.406	0.569
Total kg/sec	5095.64	105.49	5422.55	5398.57	226.62	146.93	237.71	39.15	43.35

TABLE 5-29 (continued)

Stream No.	18	19	20	21	22	23	24	25
Component Description	Solvent Effluent From Solvent Dryer	Solvent From Air Dryer	Condensate to CO ₂ Removal System	Condensate to CO ₂ Absorber	Condensate to CO ₂ Stripper	Condensate From CO ₂ Absorber	Condensate From CO ₂ Stripper	Air to CO ₂ Stripper
Composition (kg-mole/sec)								
H ₂	--	--	--	--	--	--	--	--
CH ₄	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--
CO ₂	--	--	--	--	--	--	--	--
Air	--	--	--	--	--	--	--	2.290
H ₂ O	0.019	0.070	0.602	0.045	0.350	0.019	0.093	0.068
NMP	0.595	0.197	--	--	--	--	0.005	--
Total kg-mole/sec	0.614	0.267	0.602	0.045	0.350	0.019	0.098	2.358
Total kg/sec	59.27	20.77	10.84	0.81	6.30	0.34	2.17	67.14

TABLE 5-29 (continued)

Stream No.	26	27	28	29	30	31	32
Component Description	Methanator Effluent	Recycle Condensate	Makeup Water	Process Water to Section 600	CO ₂ to Atmosphere	Recycle Hydrogen	Product Hydrogen
Composition (kg-mole/sec)							
H ₂	26.684	--	--	--	0.259	12.807	13.877
CH ₄	2.601	--	--	--	--	1.248	1.353
CO	--	--	--	--	0.008	--	--
CO ₂	--	--	--	--	6.919	--	--
Air	--	--	--	--	2.290	--	--
H ₂ O	0.559	0.532	14.569	23.125	0.764	0.013	0.014
NMP	--	--	--	--	--	--	--
Total kg-mole/sec	29.844	0.532	14.569	23.125	10.240	14.068	15.244
Total kg/sec	105.49	9.58	262.23	416.23	384.64	46.02	49.89

Carbon dioxide removal is accomplished by scrubbing with NMP (N-methyl-pyrrolidone) by the Purisol process. After drying, the process gas is compressed to 10.5 MN/m^2 (1515 psia) by multistage centrifugal compressors (R-701). The high-pressure gas is scrubbed with NMP in the CO_2 absorbers (A-701) to remove carbon dioxide. The rich solvent from the CO_2 absorbers is reduced in pressure to 3.00 MN/m^2 (435 psia) through a power recovery turbine. The evolved gases are rich in hydrogen and are recovered in a flash drum (M-702) for recycle. The solvent is regenerated by flashing at 0.14 MN/m^2 (20 psia) and then stripping with air in the CO_2 stripper (A-702).

A small slipstream of solvent is used to dry the compressor feed gas and the air required for stripping. A water balance is maintained in the solvent dryer (A-704) by heating and rejecting the water overhead. Gases leaving the CO_2 absorber, the CO_2 stripper, and the solvent dryer are washed with water to reduce solvent losses.

The CO_2 absorber effluent gas is fed to the methanators (L-703), where remaining carbon oxide impurities are reduced to trace levels. The methanater effluent is cooled and the condensate is recycled for treatment. A portion of the hydrogen stream is compressed to 13.78 MN/m^2 (2000 psia) to provide the hydrogen required for gasification of the coal. The product hydrogen has a purity of 91.0%.

5.2.5. Case V, Process Temperature = 1367°K (2000°F)

A summary of the plant performance is given in Table 5-30, and additional details on the power requirements and cooling water requirements are given in Tables 5-31 and 5-32. Overall stream flows are given in Figs. 5-17 and 5-18 and Tables 5-33 and 5-34.

The required changes to the base case were identical to the changes for Case IV [process temperature = 1256°K (1800°F)].

TABLE 5-30
OVERALL PERFORMANCE SUMMARY, CASE V, PROCESS TEMPERATURE = 1367°K (2000°F)

	Quantity at 100% Capacity	Gross Heat Value		% of Total Input
		(MW)	(10 ⁹ Btu/hr)	
Input				
Coal feed	220.21 kg/sec (20,970 tpd)	6382	21.79	68.0
Makeup water	287.55 kg/sec (4559 gpm)			
HTGR	3000 MW	3000	10.24	32.0
Total in		9382	32.03	100.0
Product				
Hydrogen	17.030 kg-mole/sec (1165 MMscfd) 93.8 mol % ^(a)	5500	18.78	58.6
Aromatics	22.32 kg/sec (12,150 tpd)	946	3.23	10.1
Power	7 MW	7	0.02	0.1
Subtotal		6453	22.03	68.8
Waste Heat				
Char	40.24 kg/sec (3832 tpd)	747	2.55	8.0
To cooling water		1936	6.61	20.6
Other		246	0.84	2.6
Subtotal		2929	10.00	31.2
Total Out		9382	32.03	100.0

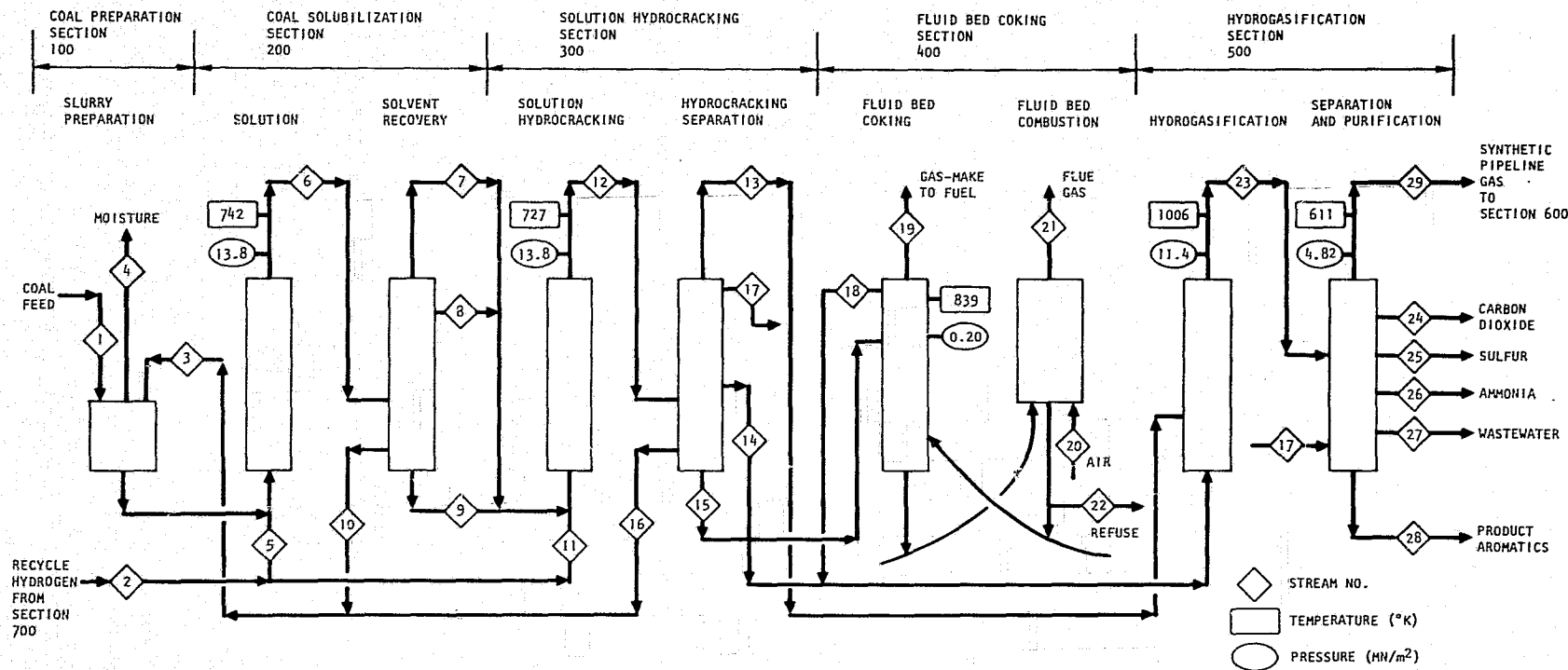
^(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

TABLE 5-31
POWER REQUIREMENTS, CASE V,
PROCESS TEMPERATURE = 1367°K (2000°F)

	Power (MW)
Section 600	
Power produced	
Circulator turbine	100
High-pressure turbine	59
Low-pressure turbine	252
Subtotal	411
Power consumed	
Helium circulator	(100)
Pumps	(14)
Subtotal	(114)
Net power produced, Section 600	297
Section 700	
Power produced	34
Power consumed	
Hydrogen compression	(215)
Pumps	(81)
Other	--
Subtotal	(296)
Net power consumed, Section 700	(262)
Net power consumed (Sections 100 to 500)	(28)
Total net power produced (Sections 100 to 700)	7

TABLE 5-32
COOLING WATER REQUIREMENTS, CASE V,
PROCESS TEMPERATURE = 1367°K (2000°F)

Item No.	Description	Heat Duty	
		MW	10 ⁹ Btu/hr
Sections 600, 700			
T-604	Power steam condenser	472	1.613
T-701	Compressor precooler	414	1.414
T-703	Absorber precooler	98	0.333
T-705	Compressor precooler	86	0.293
R-701	Compressor intercooler	98	0.333
Total Sections 600 to 700		1168 MW	(3986 MMBtu/hr)
Total Sections 100 to 500		769 MW	(2627 MMBtu/hr)
Total heat duty		1937 MW	(6613 MMBtu/hr)



OVERALL STREAM FLOWS SECTIONS 100-500
 COAL GASIFICATION, CASE V
 PROCESS TEMPERATURE = 1367°K (2000°F)

Fig. 5-17. Flow diagram, Sections 100-500, Case V

TABLE 5-33
HYDROGASIFICATION OF COAL LIQUIDS OVERALL PLANT FLOWS, CASE V, PROCESS TEMPERATURE = 1367°K (2000°F)

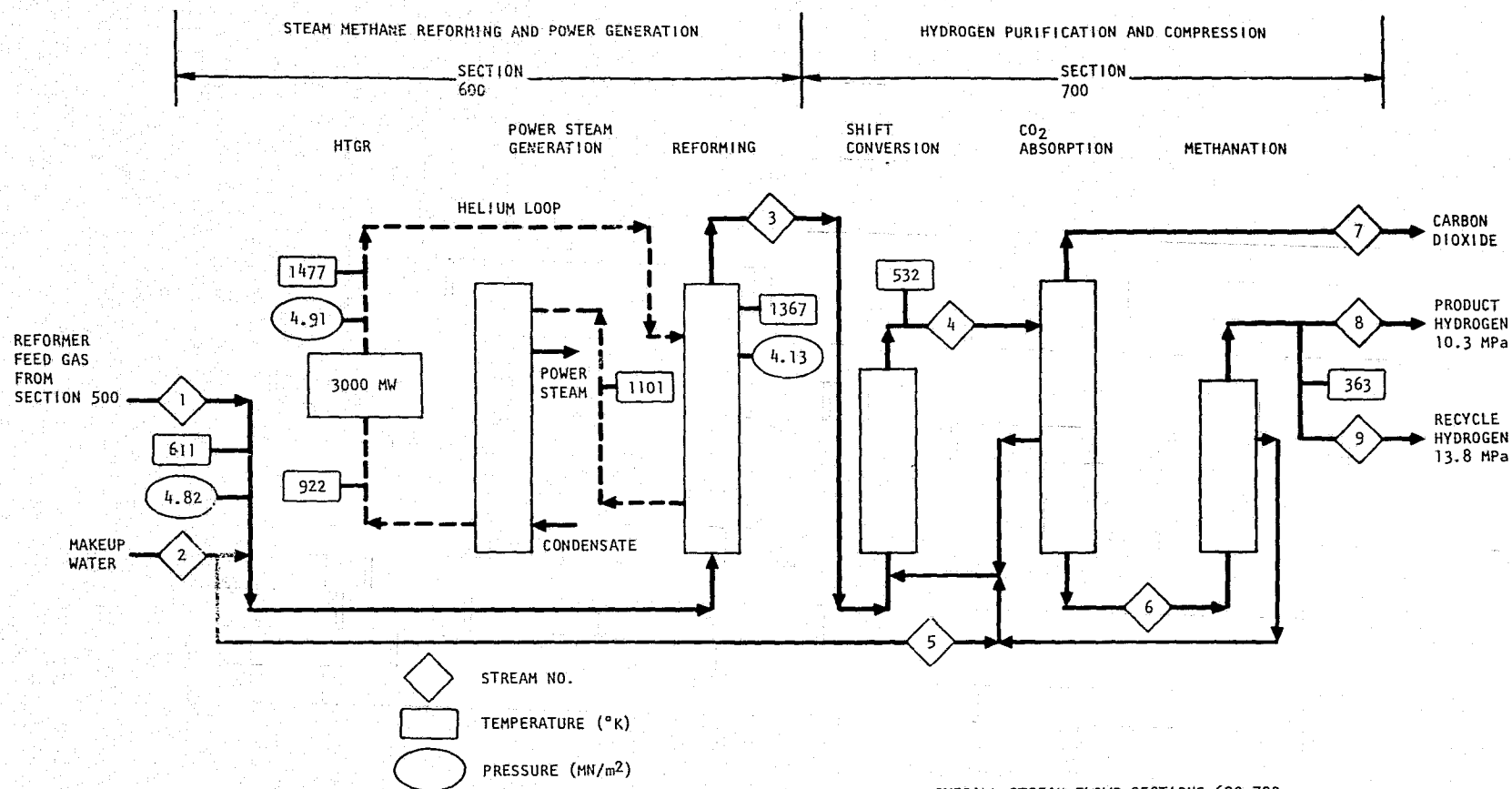
Stream No.	1	2	3	4	5	6	7	8	9	10
Component Description	Coal Feed	Recycle Hydrogen	Recycle Solvent	Moisture	Hydrogen to Coal Solution	Coal Solution Effluent	Hydrocarbon Gases	Distillate	Coal Liquids	Recycle Solvent
Composition (kg/sec)										
H ₂	--	26.95	--	--	13.49	11.03	11.03	--	--	--
C ₁ -C ₄	--	14.09	--	--	7.05	19.09	19.09	--	--	--
C ₅ -465°K	--	--	--	--	--	11.50	--	11.50	--	--
465°K-650°K	--	--	--	--	--	573.26	--	--	472.30	--
650°K-810°K	--	--	408.55	--	--		--	--		100.96
>810°K	--	--	--	--	--		--	--		--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	220.21	--	--	--	--	--	--	--	--	--
H ₂ O	--	0.23	--	11.42	0.12	23.12	23.12	--	--	--
H ₂ S	--	--	--	--	--			--	--	--
NH ₃	--	--	--	--	--			--	--	--
CO ₂	--	--	--	--	--			--	--	--
Air	--	--	--	--	--			--	--	--
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	220.21	41.27	408.55	11.42	20.66	638.00	53.24	-11.50	472.30	100.96

TABLE 5-33 (continued)

Stream No.	11	12	13	14	15	16	17	18	19	20
Component Description	Hydrogen to Solution Hydro-cracking	Solution Hydro-cracker Effluent	Hydro-cracking Product Gas	Distillate	Heavy Coal Liquids	Recycle Solvent	Acid Gases	Coal Liquids	Gas-Make to Fuel	Combustion Air
Composition (kg/sec)										
H ₂	13.46	14.37	14.37	--	--	--	--	--	} 0.80	--
C ₁ -C ₄	7.04	49.25	49.25	--	--	--	--	--		--
C ₅ -465°K	--	} 91.40	--	} 91.40	--	--	--	--		--
465°K-650°K	--		--		--	--	--	--	--	--
650°K-810°K	--	} 372.84	--	--	} 65.25	307.59	--	21.59	--	--
>810°K	--		--	--		--	--	--	--	--
Coke (inc ash)	--	--	--	--	--	--	--	--	--	--
Coal	--	--	--	--	--	--	--	--	--	--
H ₂ O	0.11	} 29.79	} 9.49	--	--	--	} 20.30	--	--	--
H ₂ S	--			--	--	--		--	--	--
NH ₃	--			--	--	--		--	--	--
CO ₂	--			--	--	--		--	--	--
Air	--	--	--	--	--	--	--	--	--	23.20
Flue Gas	--	--	--	--	--	--	--	--	--	--
Total kg/sec	20.61	557.65	73.11	91.40	65.25	307.59	20.30	21.59	0.80	23.20

TABLE 5-33 (continued)

Stream No.	21	22	23	24	25	26	27	28	29	29	
Component Description	Flue Gas	Refuse	Hydro-gasifier Effluent	Carbon Dioxide	Sulfur	Ammonia	Waste-water	Product Aromatics	Reformer Feed Gas	Reformer Feed Composition (kg-mole/sec)	
Composition (kg/sec)											
H ₂	--	--	2.59	--	--	--	--	--	2.59	H ₂	1.288
C ₁ -C ₄	--	--	148.57	--	--	--	--	--	148.57	CH ₄	5.819
C ₅ -465°K	--	--	22.32	--	--	--	--	22.32	--	C ₂ H ₆	1.328
465°K-650°K	--	--	--	--	--	--	--	--	--	C ₃ H ₈	0.173
650°K-810°K	--	--	--	--	--	--	--	--	--	C ₄ H ₁₀	0.134
>810°K	--	--	--	--	--	--	--	--	--		8.742
Coke (inc ash)	--	40.24	--	--	--	--	--	--	--	Total	
Coal	--	--	--	--	--	--	--	--	--	(kg-mole/sec)	
H ₂ O	--	--	12.62	--	--	--	20.23	--	--		
H ₂ S	--	--		--	5.61	--	--	--	--		
NH ₃	--	--		--	--	3.41	--	--	--		
CO ₂	--	--		3.67	--	--	--	--	--		
Air	--	--	--	--	--	--	--	--	--		
Flue Gas	25.82	--	--	--	--	--	--	--	--		
Total kg/sec	25.82	40.24	186.10	3.67	5.61	3.41	20.23	22.32	151.16		



OVERALL STREAM FLOWS SECTIONS 600-700
HYDROGEN PRODUCTION, CASE V
PROCESS TEMPERATURE = 1367°K (2000°F)

Fig. 5-18. Flow diagram, Sections 600-700, Case V

TABLE 5-34

HYDROGEN PRODUCTION AND POWER GENERATION OVERALL STREAM FLOWS, CASE V, PROCESS TEMPERATURE = 1367°K (2000°F)

Stream No.	1	2	3	4	5	6	7	8	9	Stream No.	8	9
Component Description	Reformer Feed Gas	Makeup Water	Reformer Effluent	Shift Converter Effluent	Recycle Condensate	Carbon Dioxide Absorber Effluent	Carbon Dioxide	Recycle Hydrogen to Coal Liquefaction	Product Hydrogen	Hydrogen Composition (kg/sec)	Recycle Hydrogen	Product Hydrogen
Composition (kg-mole/sec)												
H ₂	1.288	--	24.412	31.333	--	31.068	0.265	13.381	15.966	H ₂	26.95	32.16
CH ₄	5.819	--	1.398	1.398	--	1.398	--	0.879	1.049	CH ₄	14.09	16.81
C ₂ H ₆	1.328	--	--	--	--	--	--	--	--	CO	--	--
C ₃ H ₈	0.173	--	--	--	--	--	--	--	--	CO ₂	--	--
C ₄ H ₁₀	0.134	--	--	--	--	--	--	--	--	H ₂ O	0.23	0.27
CO	--	--	7.328	0.407	--	0.399	0.008	--	--	Total kg/sec	41.27	49.24
CO ₂	--	--	0.804	7.725	--	0.131	7.594	--	--			
H ₂ O	--	15.976	5.376	15.055	1.664	0.028	0.752	0.013	0.015			
Total kg-mole/sec	8.742	15.976	39.318	55.918	1.664	33.024	8.619	14.273	17.030			
kg/sec	151.16	287.55	408.76	707.54	29.95	102.41	348.20	41.27	49.24			

5.2.5.1. Hydrogen Production and Power Generation (Section 600). The detailed changes in process conditions are given in the flow sheet shown in Fig. 5-19 and the mass balance given in Table 5-35.

Hot helium at 1477°K (2200°F) carries heat from the HTGR nuclear core to the steam generators and reformers. The distribution of HTGR thermal power is as follows:

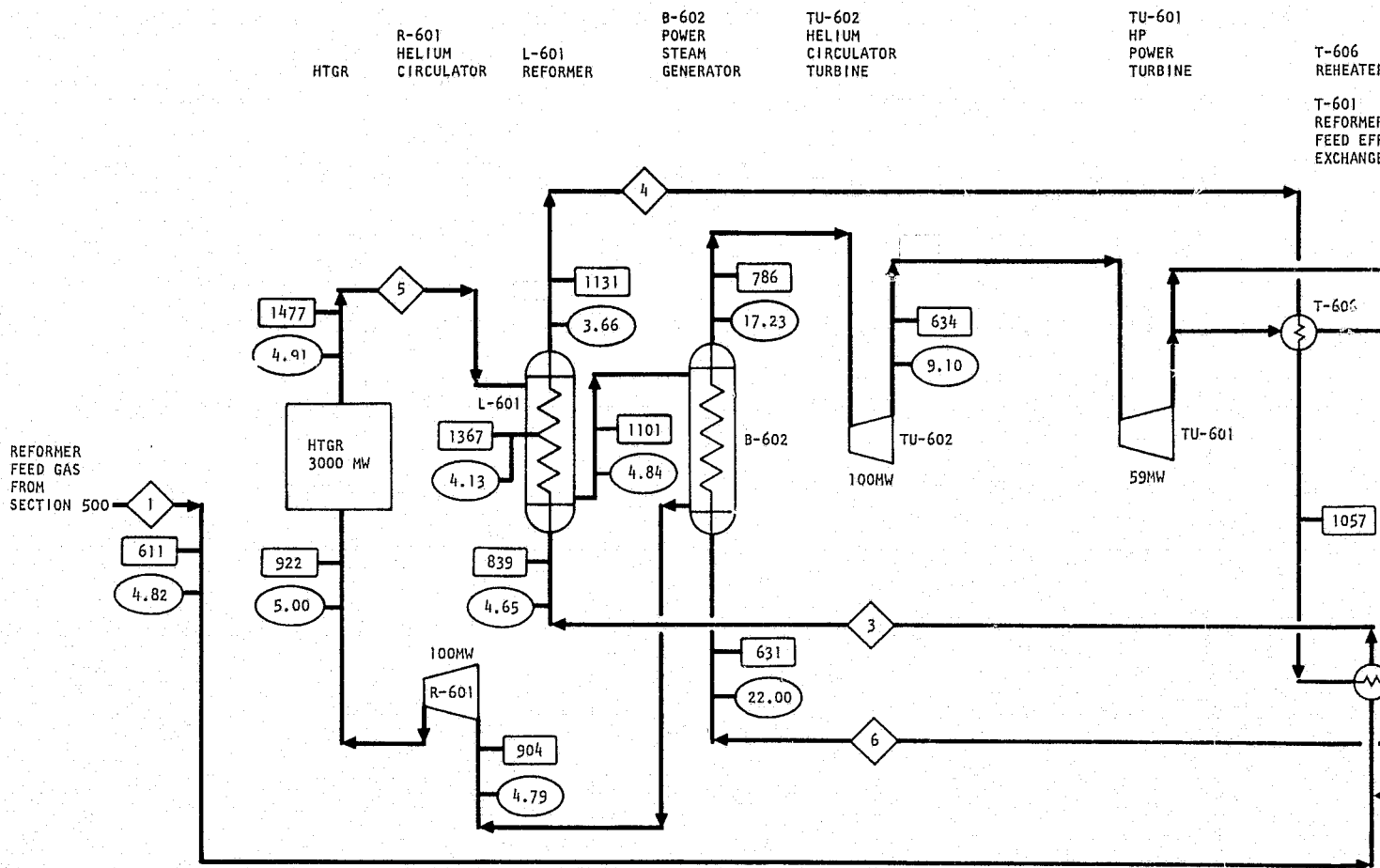
	<u>Thermal Power (MW)</u>
Reformers	2036
Steam generators	<u>1064</u>
Subtotal	3100
Helium circulators	<u>-100</u>
Total HTGR power	3000

The process steps are identical to those in Case IV [process temperature = 1256°K (1800°F)].

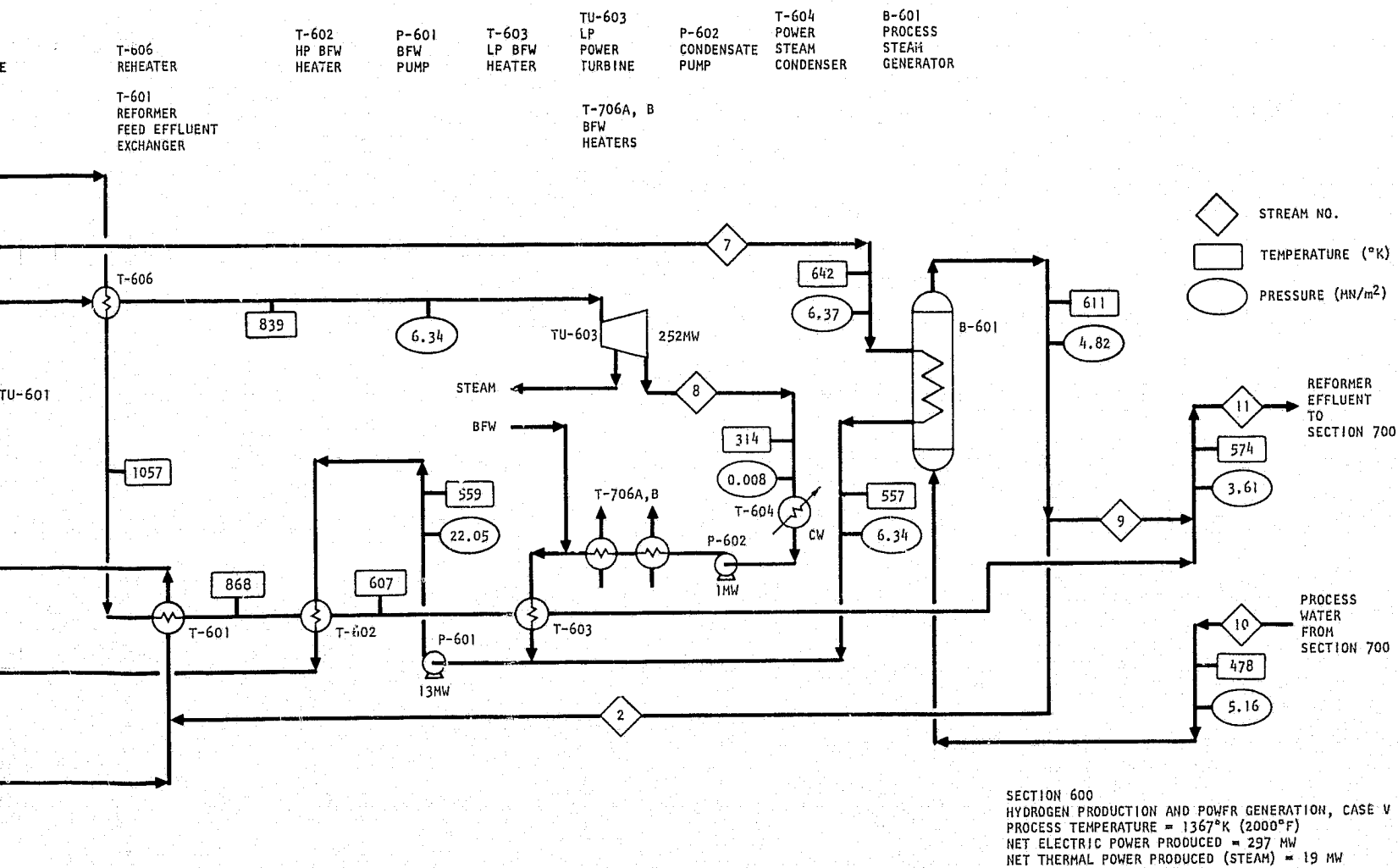
5.2.5.2. Hydrogen Purification and Compression (Section 700). The process steps are identical to those for Case IV [process temperature = 1256°K (1800°F)]. Detailed changes in process conditions are given in the flow sheet shown in Fig. 5-20 and the mass balance given in Table 5-36. Catalyst bed discharge conditions are 1256°K (1800°F) and 2.76 MN/m² (400 psia), and the process gas exits through integral return tubes leaving the reformer at 1131°K (1575°F). Hydrogen purity after purification is 93.8%.

REFERENCE

- 5-1. "High-Temperature Nuclear Heat Source Study," USAEC Report GA-A13158, General Atomic, December 30, 1974.



FOLDOUT FRAME



FOLDOUT FRAME 2

Fig. 5-19. Flow diagram, Section 600, Case V

TABLE 5-35
HYDROGEN PRODUCTION AND POWER GENERATION STREAM FLOWS, CASE V, PROCESS TEMPERATURE = 1367°K (2000°F)

Stream No.	1	2	3	4	5	6	7	8	9	10	11
Component Description	Reformer Feed Gas	Reformer Steam	Reformer Feed	Reformer Effluent	Helium Heat Transfer Loop	Power Cycle BFW	Steam to Process Steam Generator	Steam From LP Turbine	Steam to 1st-Stage Shift Converter	Process BFW	1st-Stage Shift Converter Feed
Composition (kg-mole/sec)											
He	--	--	--	--	259.715	--	--	--	--	--	--
H ₂	1.288	--	1.288	24.412	--	--	--	--	--	--	24.412
CH ₄	5.819	--	5.819	1.398	--	--	--	--	--	--	1.398
C ₂ H ₆	1.328	--	1.328	--	--	--	--	--	--	--	--
C ₃ H ₈	0.173	--	0.173	--	--	--	--	--	--	--	--
C ₄ H ₁₀	0.134	--	0.134	--	--	--	--	--	--	--	--
CO	--	--	--	7.328	--	--	--	--	--	--	7.328
CO ₂	--	--	--	0.804	--	--	--	--	--	--	0.804
H ₂ O	--	14.312	14.312	5.376	--	37.830	25.580	11.780	7.396	21.708	12.772
Total kg-mole/sec	8.742	14.312	23.054	39.318	259.715	37.830	25.580	11.780	7.396	21.708	46.714
Total kg/sec	151.16	257.60	408.76	408.76	1038.60	680.90	460.41	212.03	133.12	390.72	541.88

L-701
1ST-STAGE
SHIFT
CONVERTER

L-702
2ND-STAGE
SHIFT
CONVERTER

T-706A,B
BFW
HEATERS

T-702A,B
BFW
HEATERS

P-703
BFW
PUMP

H-701
KO
DRUM

H-702
FLASH
DRUM

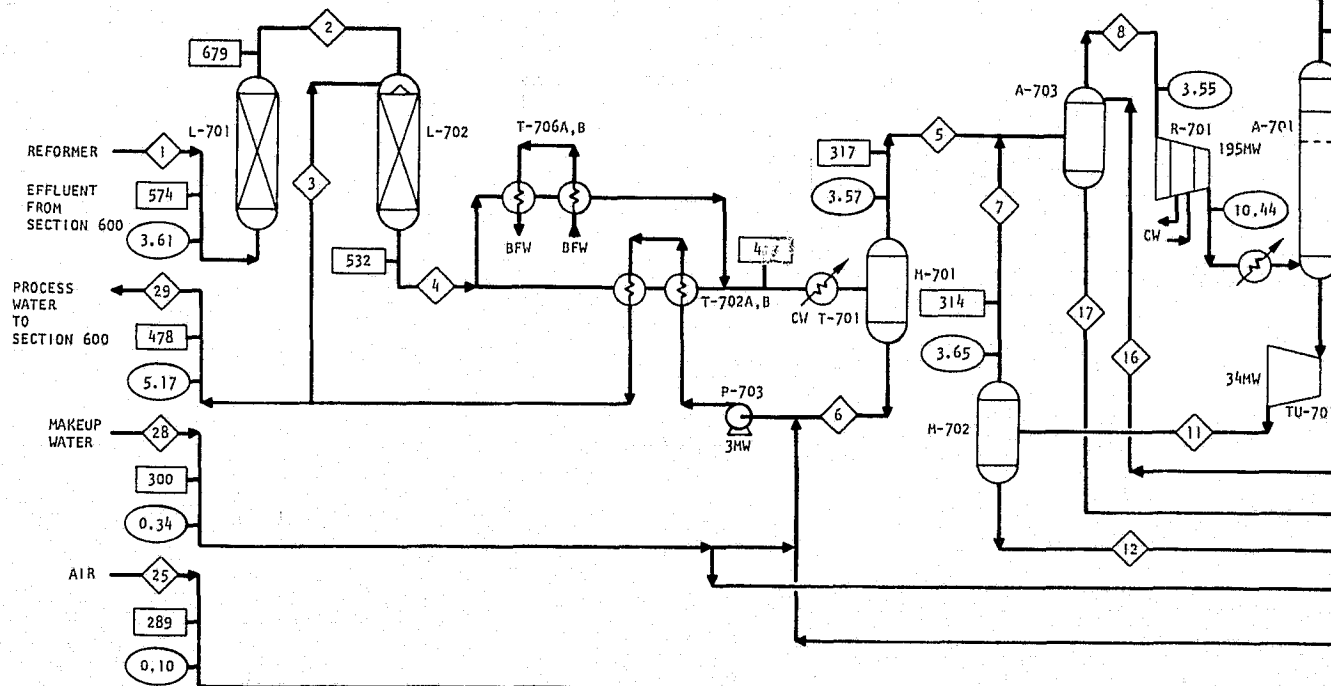
A-703
PROCESS
GAS DRYER

T-703
ABSORBER
PRECOOLER

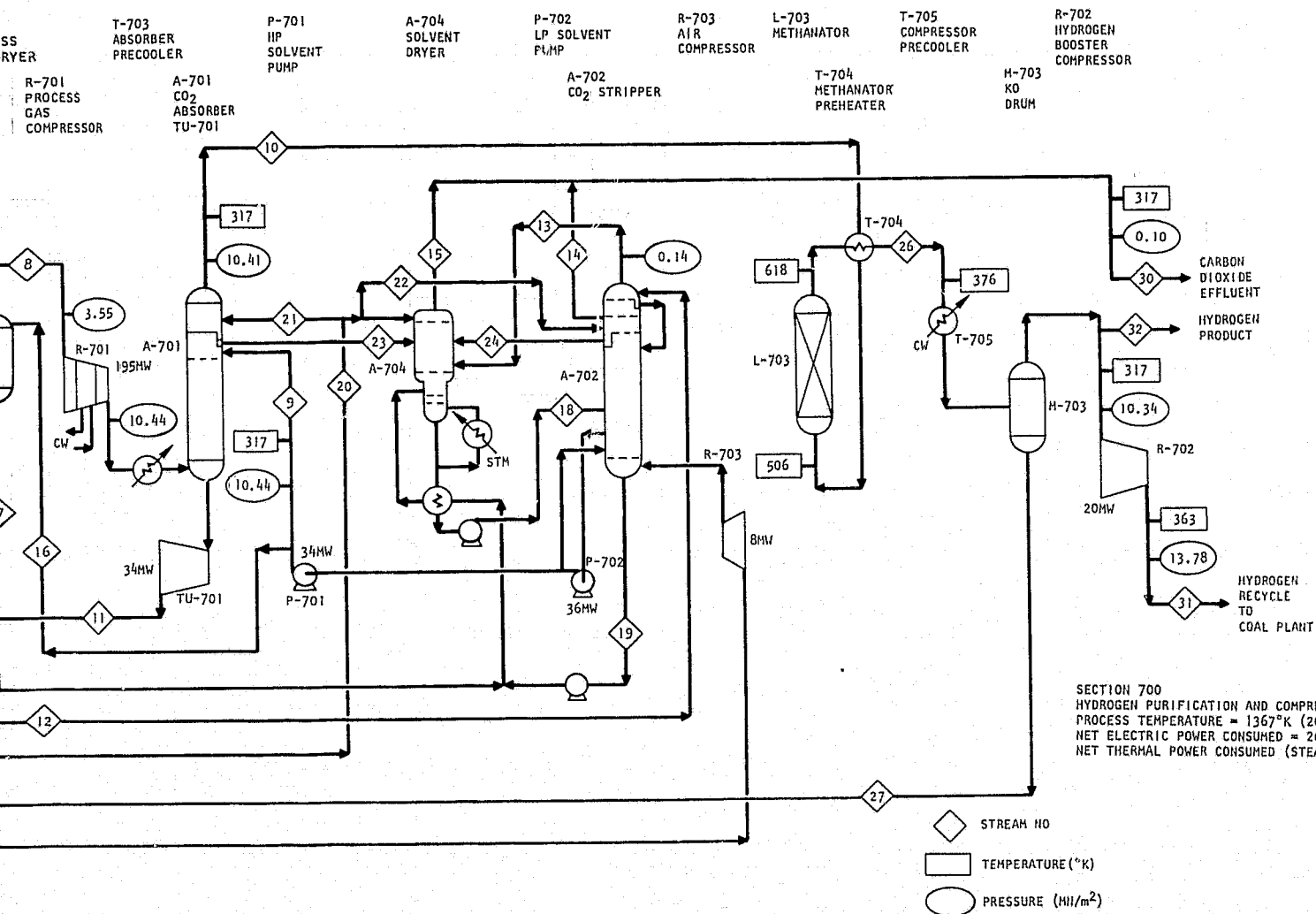
T-701
COMPRESSOR
PRECOOLER

R-701
PROCESS
GAS
COMPRESSOR

A-701
CO₂
ABSORBER
TU-701



FOLDOUT FRAME



FOLDOUT FRAME 2

Fig. 5-20. Flow diagram, Section 700, Case V

TABLE 5-36
HYDROGEN PURIFICATION AND COMPRESSION STREAM FLOWS, CASE V, PROCESS TEMPERATURE = 1367°K (2000°F)

Stream No.	1	2	3	4	5	6	7	8
Component Description	1st-Stage Shift Converter Feed	1st-Stage Shift Converter Effluent	Condensate to 2nd-Stage Shift Converter	2nd-Stage Shift Converter Effluent	Feed Gas to CO ₂ Removal System	Recycle Condensate	Recycle Hydrogen	CO ₂ Absorber Feed
Composition (kg-mole/sec)								
H ₂	24.412	28.782	--	31.333	31.333	--	0.717	32.048
CH ₄	1.398	1.398	--	1.398	1.398	--	--	1.398
CO	7.328	2.958	--	0.407	0.407	--	0.011	0.418
CO ₂	6.804	5.174	--	7.725	7.725	--	0.451	8.140
Air	--	--	--	--	--	--	--	--
H ₂ O	12.772	8.402	9.204	15.055	0.100	14.955	--	0.010
NMP	--	--	--	--	--	--	--	--
Total kg-mole/sec	46.714	46.714	9.204	55.918	40.963	14.955	1.179	42.014
Total kg/sec	541.88	541.88	165.66	707.54	438.37	269.17	21.58	456.75

TABLE 5-36 (continued)

Stream No.	9	10	11	12	13	14	15	16	17
Component Description	Lean Solvent to Absorber	CO ₂ Absorber Effluent	Rich Solvent to Flash Drum	Solvent to CO ₂ Stripper	LP Flash Gas	CO ₂ Stripper Effluent Gas	Solvent Dryer Effluent Gas	Solvent to Dryer	Solvent Effluent From Dryer
Composition (kg-mole/sec)									
H ₂	Trace	31.068	0.980	0.263	0.263	--	0.265	--	0.002
CH ₄	--	1.398	--	--	--	--	--	--	--
CO	Trace	0.399	0.019	0.008	0.008	--	0.008	--	--
CO ₂	0.054	0.131	8.063	7.612	5.649	1.909	5.685	--	0.036
Air	--	--	--	--	--	2.520	--	--	--
H ₂ O	1.765	0.028	1.775	1.775	0.009	0.281	0.546	0.009	0.099
NMP	55.402	--	55.401	55.401	--	--	--	0.283	0.283
Total kg-mole/sec	57.221	33.024	66.238	65.059	5.929	4.710	6.504	0.292	0.420
Total kg/sec	5521.27	102.41	5876.01	5854.43	249.30	161.54	260.55	28.19	31.39

TABLE 5-36 (continued)

Stream No.	18	19	20	21	22	23	24	25
Component Description	Solvent Effluent From Solvent Dryer	Solvent From Air Dryer	Condensate to CO ₂ Removal System	Condensate to CO ₂ Absorber	Condensate to CO ₂ Stripper	Condensate From CO ₂ Absorber	Condensate From CO ₂ Stripper	Air to CO ₂ Stripper
Composition (kg-mole/sec)								
H ₂	--	--	--	--	--	--	--	--
CH ₄	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--
CO ₂	--	--	--	--	--	--	--	--
Air	--	--	--	--	--	--	--	2.520
H ₂ O	0.016	0.075	0.680	0.048	0.373	0.020	0.100	0.075
NMP	0.504	0.214	--	--	--	0.001	0.006	--
Total kg-mole/sec	0.520	0.289	0.680	0.048	0.373	0.021	0.106	2.595
Total kg/sec	50.21	22.55	12.24	0.86	6.71	0.46	2.39	73.89

TABLE 5-36 (continued)

Stream No.	26	27	28	29	30	31	32
Component Description	Methanator Effluent	Recycle Condensate	Makeup Water	Process Water to Section 600	CO ₂ to Atmosphere	Recycle Hydrogen	Hydrogen Product
Composition (kg-mole/sec)							
H ₂	29.347	--	--	--	0.265	13.381	15.966
CH ₄	1.928	--	--	--	--	0.879	1.049
CO	--	--	--	--	0.008	--	--
CO ₂	--	--	--	--	7.594	--	--
Air	--	--	--	--	2.520	--	--
H ₂ O	0.689	0.661	15.976	21.708	0.827	0.013	0.015
NMP	--	--	--	--	--	--	--
Total kg-mole/sec	31.964	0.661	15.976	21.708	11.214	14.273	17.030
Total kg/sec	102.41	11.90	287.55	390.72	422.09	41.27	49.24

N76-15583

6. CAPITAL COST ESTIMATE

6.1. REACTOR PLANT COSTS

The capital cost estimate for the nuclear process heat source (NPHS) plant was made by: (1) using costs from the current commercial HTGR for electricity production as a base for items that are essentially the same and (2) development of new estimates for modified or new equipment that is specifically for the process heat application. This procedure was used for a study by GA (Ref. 6-1) completed in December 1974, and the results of that work form the basis for the estimate used here. Some revisions have been made to reflect a typical site as used in NUS-531. Costs are shown in Table 6-1.

A balance of reactor plant (BORP) was developed by United Engineers and Constructors as a follow on to that study (Ref. 6-2). This estimate applies to the 1144°K (1600°F) process temperature. These costs are not expected to change significantly for higher process temperatures of up to 1367°K (2000°F). For process temperature of 922°K (1200°F) and 1033°K (1400°F), adjustments to the balance of reactor plant estimates were made. These adjustments, identical for both cases and listed below, reflecting changes in PCRV dimensions, fuel cycle, number of helium loops, and helium inventory, were made to Table 6-2.

<u>Account</u>	<u>Description</u>	<u>Cost Adjustment (\$ Million)</u>
212	Reactor containment building	+2.3
215	Reactor service building	-0.8
222	Main heat transfer loop	+0.6
226	Other reactor plant equipment	+0.6
263	Reactor coolant	+0.1
91	Engineering and construction costs	+2.7
	Total cost adjustment	+5.5

TABLE 6-1
COSTS FOR NUCLEAR PROCESS HEAT SOURCE
(Thousands of Dollars, July 1974)

Standard NSS List ^(a)	Differential Costs (Base Standard NSS)				
	NPS				
	1200°F Case	1400°F Case	1600°F Case	1800°F Case	2000°F Case
221 Reactor Equipment	33,078.9	37,310.7	27,875.2	34,886.5	39,311.4
PCRV liners, penetrations, and closures (L,M)					
PCRV thermal barrier (E,L,M)					
PCRV pressure relief system (E)					
Control rod drives and reserve shutdown assembly and material (E)					
Reactor core assembly (E)					
PCRV support structure (L,M)					
Prestressed concrete reactor vessel (L,M)					
Precast panels (L,M)					
Prestressing system (L,M)					
Wire-winding machine (lease) (M)					
222 Main Heat Transfer and Transport Systems	30,292.8	26,940.1	53,938.7	58,302.1	55,463.4
Main helium circulator (E)					
Steam generator module (E)					
Main circulator speed regulating equipment (E)					

TABLE 6-1 (Continued)

Standard NSS List ^(a)	Differential Costs (Base Standard NSS)				
	NPS				
	1200°F Case	1400°F Case	1600°F Case	1800°F Case	2000°F Case
Main helium circulator service system (E)					
Reformer module ^(b) (E)	36,949.3	42,423.4	64,047.3	66,627.8	66,627.8
223 Safeguards Cooling Systems	--	--	16.3	17.1	17.6
Auxiliary helium circulators (E)					
Core auxiliary heat exchanger module (E)					
Auxiliary helium circulator service system (E)					
224 Radioactive Waste Treatment and Disposal	567.4	639.8	984.8	959.8	1,188.8
Radioactive gas recovery system (E)					
Radioactive gas waste system (E)					
Radioactive liquid waste system (E)					
Radioactive solid waste system (E)					
225 Nuclear Fuel Handling and Storage Equipment	--	--	708.3	737.3	770.9
Fuel handling tools and equipment (E)					
Fuel and reflector storage wells (E)					
226 Other Reactor Plant Equipment	2,699.7	2,699.7	3,702.7	3,853.8	7,763.4
Helium purification system (E)					

TABLE 6-1 (Continued)

Standard NSS List ^(a)	Differential Costs (Base Standard NSS)				
	NPS				
	1200°F Case	1400°F Case	1600°F Case	1800°F Case	2000°F Case
227 Instrumentation and Control	79.2	79.2	72.7	75.5	78.9
PCRV instrumentation and electrical equipment (E)					
FHM control station (E)					
PCRV seal and purge system (E)					
Helium purification system instrumentation equipment (E)					
Nuclear and in-core instrumentation equipment (E)					
Analytical instrumentation system equipment (E)					
Reserve shutdown actuation system equipment and instrumentation (E)					
Rod control system instrumentation equipment (E)					
Variable orifice control system instrumentation equipment (E)					
Primary coolant instrumentation equipment (E)					
Main circulator signal control board (E)					
PPS instrumentation equipment (E)					
Overall plant control system instrumentation equipment (E)					

TABLE 6-1 (Continued)

Standard NSS List ^(a)	Differential Costs (Base Standard NSS)				
	NPS				
	1200°F Case	1400°F Case	1600°F Case	1800°F Case	2000°F Cast
Radioactive monitoring system instrumentation equipment (E)					
Operational protection system (E)					
DAP - dual computer - NSS (E)					
910 Engineering Construction Management and Field Supervision (L,M)					
<u>Other</u>					
Operator training program (L,M)					
Triaxial seismic					
Special service and installation equipment (E)					
NPS Total Delta	66,518.0	67,669.5	88,586.4 ^(c)	100,173.1 ^(c)	105,997.6 ^(c)
NSS Base Cost ^(d)	133,000.0	133,000.0	133,000.0	133,000.0	133,000.0
Total Cost Nuclear Process Heat System	199,500.0	200,700.0	221,600.0	233,200.0	239,000.0

(a) E = equipment
L = labor
M = materials.

(b) Reformer module cost differentials are included in account 222 total differential.

(c) An additional contingency to cover added risks was added: \$24,000,000 at 1600°F, \$30,000,000 at 1800°F, and \$35,000,000 at 2000°F.

(d) Basis Ref. (1), corrected for typical site and conditions.

TABLE 6-2
BALANCE OF REACTOR PLANT COSTS
1144°K (1600°F) PROCESS TEMPERATURE
(July, 1974)

Account	Description	Costs (Thousands of Dollars)		
		BOP Materials and Equipment	BOP Labor	Total
20	<u>Land and Land Rights</u>			
201	Land and privilege acquisition	1,000	--	1,000
Total 20		1,000	--	1,000
21	<u>Structures and Improvements</u>			
211	Yard work	1,203	1,424	2,627
212	Reactor containment building	12,247	21,564	33,811
214	Intake structure	285	127	412
215	Reactor service building	6,000	11,255	17,255
218A	Control building	1,900	4,146	6,046
218B	Diesel generator building	1,628	3,512	5,140
218C	Administration building	1,200	1,372	2,572
218D	Auxiliaries building	363	485	848
218E	Helium storage building	90	71	161
Total 21		24,916	43,956	68,872
22	<u>Reactor Plant Equipment</u>			
221	Reactor equipment	104	578	682
222	Main heat transfer loop	318	1,146	1,464
223	Safeguards cooling system	409	259	668
224	Radioactive waste treatment and disposal	239	884	1,123
225	Nuclear fuel handling and storage	254	640	894
226	Other reactor plant equipment	9,968	6,998	16,966
227	Instrumentation and control	2,356	1,169	3,525
Total 22		13,648	11,674	25,322

TABLE 6-2 (Continued)

Account	Description	Costs (Thousands of Dollars)		
		BOP Materials and Equipment	BOP Labor	Total
24	<u>Electric Plant Equipment</u>			
241	Switchgear	2,131	381	2,512
242	Station Service Equipment	4,484	808	5,292
243	Switchboards	361	146	507
244	Protective Equipment	190	328	518
245	Electrical Structures and Wiring Containers	1,429	5,615	7,044
246	Power and Control Wiring	4,886	5,838	10,724
Total 24		13,481	13,116	26,597
25	<u>Miscellaneous Plant Equipment</u>			
251	Transportation and Lifting Equipment	1,121	278	1,399
252	Air, Hydraulic, Water and Steam Service Systems	3,165	3,643	6,808
253	Communications Equipment	74	107	181
254	Furnishings and Fixtures	345	41	386
Total 25		4,705	4,069	8,774
26	<u>Special Materials</u>			
263	Reactor Coolant (and initial storage)	254	--	254
269	Initial Catalyst Filling	0	20	20
Total 26		254	20	20
Total 21, 22, 23, 24, 25, & 26		57,004	72,835	129,839

TABLE 6-2 (Continued)

Account	Description	Costs (Thousands of Dollars)		
		BOP Materials and Equipment	BOP Labor	Total
91	<u>Engineering and Construction Costs</u>			
910A	Engineering services and construction management	--	63,898	63,898
911	Temporary facilities	1,800	3,375	5,175
912	Construction equipment	8,400	900	9,300
913	Construction services	5,760	4,950	10,710
Total 91		15,960	73,123	89,083
99	<u>Other Costs</u>			
	GA other costs	--	--	--
991	Operator training	1,330	--	1,330
992	Spare parts	1,756	--	1,756
993	Preliminary operations and testing	--	(Not included in estimate	--
994	Miscellaneous costs	4,788	--	4,788
995	Contingency	--	--	--
Total 99		7,874	--	7,874
Total 91 & 99		23,834	73,123	96,957
Total Plant Costs		81,838	145,958	227,796

The basis for the estimate is as follows:

1. All costs were estimated for equipment installed in the field, with allowance for freight and installation labor.
2. All field indirect costs were included.
3. Land costs, site studies, and site improvements were included in the BOP estimate.

6.2. PROCESS PLANT

The process plant capital cost estimate was developed from a conceptual design which had the following basis:

1. A clear and level site would be available adjacent to the mine.
2. Rail, water, and road transport and utilities would be convenient to the site.
3. Piling would not be required.
4. Cooling would be supplied by a wet cooling tower system.
5. Three parallel trains would be used to provide reliability and economy.
6. Only proven processes and equipment would be included.
7. All present-day effluent standards would be met.

The estimate was based on a land cost of \$1000 per acre, labor rates for eastern Oklahoma, and July 1974 prices. No allowances were made for escalation or sales and use tax.

The equipment list identified all major equipment items used to prepare the in-house cost estimate. Where differences existed between the flow sheet and the equipment list, the equipment list represented the basis for estimating. Cost component information by process system and type of equipment for each temperature is shown in Tables 6-3 through 6-7. Costs for the off-sites are shown in Table 6-8. The equipment lists for all temperatures are contained in Appendix B.

6.3. SUMMARY

The total investment required for each of the five process temperatures studied is given in Table 6-9.

REFERENCES

- 6-1. "High-Temperature Nuclear Heat Source Study," USAEC Report GA-A13158, General Atomic, December 30, 1974.
- 6-2. "A Cost Comparison of Very High Temperature Nuclear Reactors for Process Heat Applications," United Engineers & Constructors Report UE&C-AEC-750311, March 1975.

TABLE 6-3
EQUIPMENT CAPITAL COSTS (\$ MILLION), JULY 1974,
CASE I, PROCESS TEMPERATURE = 922°K (1200°F)

Equipment	Section						
	100	200	300	400	500	600	700
Vessels	--	1.10	1.90	--	1.55	--	6.14
Boilers	--	--	--	--	--	2.26	--
Furnaces	--	0.65	--	--	--	--	--
General	1.30	0.17	0.10	2.15	0.69	13.43	28.45
Reactors	--	6.16	6.27	--	3.18	--	13.20
Pumps	--	0.70	0.16	--	0.27	2.69	0.73
Compressors	--	--	0.16	--	--	--	11.09
Heat exchangers	--	0.52	0.68	--	17.67	15.48	10.80
Total ^(a)	1.30	9.30	9.27	2.15	23.36	33.86	70.41
Total installed	6.31	21.43	20.95	9.14	46.72	70.78	237.88

Total major equipment 149.65

Total sections 100 to 700 413.2

^(a) Major equipment.

TABLE 6-4
EQUIPMENT CAPITAL COSTS (\$ MILLION), JULY 1974,
CASE II, PROCESS TEMPERATURE = 1033°K (1400°F)

Equipment	Section						
	100	200	300	400	500	600	700
Vessels	--	2.06	3.10	--	3.12	--	10.06
Boilers	--	--	--	--	--	2.90	--
Furnaces	--	1.14	--	--	--	--	--
General	2.24	0.30	0.24	3.44	1.27	6.25	--
Reactors	--	11.74	11.94	--	6.03	--	15.01
Pumps	--	0.89	0.96	--	0.25	2.07	1.04
Compressors	--	--	0.24	--	--	--	14.45
Heat exchangers	--	0.97	1.27	--	32.00	35.84	16.26
Total (a)	2.24	17.10	17.75	3.44	42.67	47.06	56.82
Total installed	10.88	39.40	40.80	14.62	85.34	108.50	187.40

Total major equipment 187.08

Total sections 100 to 700 486.9

(a) Major equipment.

TABLE 6-5
EQUIPMENT CAPITAL COSTS (\$ MILLION), JULY 1974,
CASE III, PROCESS TEMPERATURE = 1144°K (1600°F)

Equipment	Section						
	100	200	300	400	500	600	700
Vessels	--	2.15	3.18	--	3.27	--	11.03
Boilers	--	--	--	--	--	2.15	--
Furnaces	--	1.19	--	--	--	--	--
General	2.37	0.32	0.19	3.47	1.39	6.10	--
Reactors	--	12.31	12.31	--	6.39	--	18.38
Pumps	--	0.96	0.19	--	0.29	2.03	1.11
Compressors	--	--	0.27	--	--	--	14.95
Heat exchangers	--	0.98	1.35	--	35.81	31.82	16.98
Total ^(a)	2.37	17.91	17.49	3.47	47.15	42.10	62.45
Total installed	11.51	41.24	39.49	14.77	94.30	89.60	209.97

Total major equipment 192.94

Total sections 100 to 700 500.9

(a) Major equipment.

TABLE 6-6
EQUIPMENT CAPITAL COSTS (\$ MILLION), JULY 1974,
CASE IV, PROCESS TEMPERATURE = 1256°K (1800°F)

Equipment	Section						
	100	200	300	400	500	600	700
Vessels	--	2.37	3.48	--	3.58	--	20.90
Boilers	--	--	--	--	--	1.88	--
Furnaces	--	1.24	--	--	--	--	--
General	2.72	0.37	0.23	3.64	1.49	6.03	--
Reactors	--	13.26	16.22	--	6.89	--	30.51
Pumps	--	0.98	0.20	--	0.54	1.88	12.10
Compressors	--	--	0.27	--	--	--	15.31
Heat exchangers	--	1.05	1.47	--	37.56	9.50	4.31
Total ^(a)	2.72	19.27	21.87	3.64	50.06	19.29	83.13
Total installed	13.21	44.39	49.34	15.47	100.12	40.57	279.50

Total major equipment 199.98

Total sections 100 to 700 542.6

^(a) Major equipment.

TABLE 6-7
EQUIPMENT CAPITAL COSTS (\$ MILLION), JULY 1974
CASE V, PROCESS TEMPERATURE = 1367°K (2000°F)

Equipment	Section						
	100	200	300	400	500	600	700
Vessels	--	2.46	3.74	--	3.71	--	25.95
Boilers	--	--	--	--	--	1.70	--
Furnaces	--	1.27	--	--	--	--	--
General	2.80	0.40	0.25	3.74	1.55	7.97	--
Reactors	--	13.92	17.13	--	7.19	--	26.90
Pumps	--	0.99	0.20	--	0.55	1.84	13.46
Compressors	--	--	0.27	--	--	--	14.08
Heat exchangers	--	1.09	1.52	--	39.22	9.52	4.22
Total ^(a)	2.80	20.13	23.11	3.74	52.22	21.03	84.61
Total installed	13.60	46.38	52.18	15.89	104.44	44.02	284.49
Total major equipment					207.64		
Total sections 100 to 700					561.0		

(a) Major equipment.

TABLE 6-8
CAPITAL COSTS (\$ MILLION), ^(a) OFF-SITE FACILITIES

Section	Title	Temperature				
		1200	1400	1600	1800	2000
2000	Coal receiving and handling	2.61	3.82	3.93	4.11	4.22
2100	Water treating	1.84	2.53	2.67	2.83	2.99
2200	Ash handling ^(b)	--	--	--	--	--
2300	Sulfur recovery	5.33	7.81	8.04	8.41	8.63
2400	Cooling water system	16.30	16.19	15.24	15.09	14.55
2500	Flare and blowdown	1.34	1.81	1.92	2.01	2.15
2600	Buildings and off-site facilities	8.00	8.00	8.00	8.00	8.00
2700	Firefighting	2.34	2.34	2.34	2.34	2.34
2800	Fuel gas	0.44	0.49	0.50	0.51	0.52
2900	Tanks and storage	2.15	4.06	4.25	4.59	4.79
3000	Compressed air	0.44	0.49	0.50	0.51	0.52
3100	Site development	4.90	4.90	4.90	4.90	4.90
3200	Electrical	15.18	6.60	6.60	6.60	6.60
3300	Steam	0.40	0.76	0.79	0.86	0.90
	Total above (field costs)	61.27	59.80	59.68	60.76	61.11
	Total installed costs	75.23	73.43	73.51	74.61	75.04

(a) As of July 1974.

(b) The ash handling costs have been implemented in Section 400.

TABLE 6-9
TOTAL APPROXIMATE INVESTMENT FOR NUCLEAR COAL GASIFICATION PLANT^(a)
(Millions of Dollars)

Section	Title	Process Temperature				
		922°K (1200°F)	1033°K (1400°F)	1144°K (1600°F)	1256°K (1800°F)	1367°K (2000°F)
100	Coal preparation	6.3	10.9	11.5	13.2	13.6
200	Coal solution	21.4	39.4	41.2	44.4	46.4
300, 350	Solution hydrocracking	21.0	40.8	39.5	49.3	52.2
400	Fluid bed coking	9.1	14.6	14.8	15.5	15.9
500, 550	Hydrogasification	46.7	85.3	94.3	100.1	104.4
600	Power generator	70.8	108.5	89.6	40.6	44.0
700	Hydrogen production	237.9	187.4	210.0	279.5	284.5
1000	HTGR system reformer	432.8	434.0	449.4	461.0	466.8
2000, 3000	Off-site facilities	75.2	73.4	73.5	74.6	75.0
	Subtotal	921.2	994.3	1023.8	1078.2	1102.8
	Interest during construction	180.1	192.4	198.2	208.1	212.5
	Total fixed investment	1101.3	1186.7	1222.0	1286.3	1315.3

^(a) Based on July 1974 figures.

7. ECONOMICS

The selling price of hydrogen is given in Tables 7-1 through 7-5 for each of the process temperatures studied. The estimated selling price was based on the capital costs given in Section 6 and the operating and maintenance costs given in Tables 7-6 and 7-7. In all cases, no credit was given for the methane component of hydrogen.

7.1. EFFECT OF PROCESS TEMPERATURE

The effect of process temperature on selling price of hydrogen is shown in Fig. 7-1. The selling price decreases with increasing process temperature over the entire range of temperatures studied [from 922°K (1200°F) to 1367°K (2000°F)]. The selling price is most sensitive to process temperatures of between 922°K (1200°F) and 1033°K (1400°F).

7.2. EFFECT OF COAL COST

The effect of coal cost on selling price is shown in Fig. 7-2. The base prices were developed for a coal cost of \$11/Kkg (\$10/st) and a credit for aromatics of \$71/Kkg (\$10/barrel). The credit for aromatics was varied in proportion to the coal cost in order to reflect its superiority over the coal as a sulfur free fuel. The cost of coal has a small but significant effect on hydrogen price. Increasing the coal cost and aromatics credit by a factor of 4 adds about \$0.08/kg to \$0.12/kg to the hydrogen price. Optimization of the process for a higher coal cost could reduce this effect significantly.

7.3. EFFECT OF NUCLEAR FUEL COST

The effect of nuclear fuel cost on hydrogen selling price is shown in Fig. 7-3. The base price estimates were for nuclear fuel costs ranging

TABLE 7-1
HYDROGEN PRICE ESTIMATE, CASE I, PROCESS TEMPERATURE = 922°K (1200°F),
HYDROGEN PRODUCED = 13.79 KG/SEC (130 MMSCFD) (a) AT 0.9 PLANT CAPACITY FACTOR

	Process Plant			HTGR			Total	
Plant investment (\$10 ⁶)								
Process units	413.2			199.5 (NPS)			--	
Off-site facilities	75.2			233.3 (BOP)			--	
Interest during construction (b)	81.2			98.9			--	
Total investment	569.6			531.7			1,101.3	
Working capital	--			--			11.7	
Total capital required	--			--			1,113.0	

	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Cost of coal charge = \$11/Kkg	7,671	30,801	--	--	--	--	--	--
Cost of nuclear fuel = \$0.4436/GJ	--	--	--	--	37,776	--	--	--
Operation and maintenance	--	23,458	--	--	4,326	--	--	--
Subtotal expenses	--	54,259	12.47	--	42,102	9.68	96,361	22.15
Byproduct credits (c)								
Aromatics = \$71/Kkg	792	20,533	--	--	--	--	--	--
Ammonia = \$38.5/Kkg NH ₃	119	1,676	--	--	--	--	--	--
Sulfur = \$10/Kkg	185	673	--	--	--	--	--	--
Char = /Kkg	1,401	--	--	--	--	--	--	--
Electric power = \$0.01/kWh	2,548,000	9,301	--	--	--	--	--	--
Subtotal credits		32,183	7.40	--	--	--	32,183	7.40
Expenses after credits	--	--	--	--	--	--	64,178	14.75

TABLE 7-1 (continued)

	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Hydrogen price (utility basis)								
Capital charge (15%)	--	--	--	--	--	--	166,950	38.38
Hydrogen price	--	--	--	--	--	--	--	53.13
Hydrogen price (industry basis)								
Capital charge (25%)	--	--	--	--	--	--	278,250	63.97
Hydrogen price ^(c)	--	--	--	--	--	--	--	78.72

(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

(b) Eight percent per year.

(c) No credit given for methane content of product.

TABLE 7-2
HYDROGEN PRICE ESTIMATE, CASE II, PROCESS TEMPERATURE = 1033°K (1400°F),
HYDROGEN PRODUCED = 20.22 KG/SEC (686 MMSCFD)^(a) AT 0.9 PLANT CAPACITY FACTOR

	Process Plant			HTGR			Total	
Plant investment (\$10 ⁶)								
Process units	486.9			200.7 (NPS)			--	
Off-site facilities	73.4			233.3 (BOP)			--	
Interest during construction ^(b)	93.2			99.2			--	
Total investment	653.5			533.2			1,186.7	
Working capital	--			--			18.1	
Total capital required	--			--			1,204.8	
	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Cost of coal charge = \$11/Kkg	14,518	58,292	--	--	--	--	--	--
Cost of nuclear fuel = \$0.4479/GJ	--	--	--	--	38,142	--	--	--
Operation and maintenance	--	28,387	--	--	4,007	--	--	--
Subtotal expenses	--	86,679	13.59	--	42,149	6.61	128,828	20.20
Byproduct credits ^(c)								
Aromatics = \$71/Kkg	1,472	38,134	--	--	--	--	--	--
Ammonia = \$38.5/Kkg NH ₃	225	3,157	--	--	--	--	--	--
Sulfur = \$10/Kkg	349	1,271	--	--	--	--	--	--
Char = /Kkg	2,653	--	--	--	--	--	--	--
Subtotal credits	--	42,562	6.67	--	--	--	42,562	6.67
Expenses after credits	--	--	--	--	--	--	86,266	13.53

TABLE 7-2 (continued)

	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Hydrogen price (utility basis)								
Capital charge (15%)	--	--	--	--	--	--	180,720	28.33
Hydrogen price	--	--	--	--	--	--	--	41.86
Hydrogen price (industry basis)								
Capital charge (25%)	--	--	--	--	--	--	301,200	47.22
Hydrogen price ^(c)	--	--	--	--	--	--	--	60.75

(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

(b) Eight percent per year.

(c) No credit given for methane content of product.

TABLE 7-3
HYDROGEN PRICE ESTIMATE, CASE III, PROCESS TEMPERATURE = 1144°K (1600°F),
HYDROGEN PRODUCED = 23.39 KG/SEC (794 MMSCFD)^(a) AT 0.9 PLANT CAPACITY FACTOR

	Process Plant			HTGR			Total	
Plant investment (\$10 ⁶)								
Process units	500.9			221.6 (NPS)			--	
Off-site facilities	73.5			227.8 (BOP)			--	
Interest during construction ^(b)	95.5			102.7			--	
Total investment	569.9			552.1			1,222.0	
Working capital	--			--			18.9	
Total capital required	--			--			1,240.9	
	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Cost of coal charge = \$11/Kkg	15,214	61,086	--	--	--	--	--	--
Cost of nuclear fuel = \$0.4929/GJ	--	--	--	--	41,957	--	--	--
Operation and maintenance	--	29,472	--	--	3,827	--	--	--
Subtotal expenses	--	90,558	12.27	--	45,785	6.21	136,342	18.48
Byproduct credits ^(c)								
Aromatics = \$71/Kkg	1,536	39,799	--	--	--	--	--	--
Ammonia = \$38.5/Kkg NH ₃	235	3,304	--	--	--	--	--	--
Sulfur = \$10/Kkg	366	1,333	--	--	--	--	--	--
Char = /Kkg	2,780	--	--	--	--	--	--	--
Subtotal credits	--	44,436	6.02	--	--	--	44,436	6.02
Expenses after credits	--	--	--	--	--	--	91,906	12.46

TABLE 7-3 (continued)

	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Hydrogen price (utility basis)								
Capital charge (15%)	--	--	--	--	--	--	186,135	25.23
Hydrogen price	--	--	--	--	--	--	--	37.69
Hydrogen price (industry basis)								
Capital charge (25%)	--	--	--	--	--	--	310,225	42.05
Hydrogen price ^(c)	--	--	--	--	--	--	--	54.51

(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

(b) Eight percent per year.

(c) No credit given for methane content of product.

TABLE 7-4
HYDROGEN PRICE ESTIMATE, CASE IV, PROCESS TEMPERATURE = 1256°K (1800°F),
HYDROGEN PRODUCED = 25.16 KG/SEC (854 MMSCFD) (a) AT 0.9 PLANT CAPACITY FACTOR

	Process Plant			HTGR			Total	
Plant investment (\$10 ⁶)								
Process units	542.6			233.2 (NPS)			---	
Off-site facilities	74.6			227.8 (BOP)			---	
Interest during construction (b)	102.7			105.4			---	
Total investment	719.9			566.4			1,286.3	
Working capital	---			---			20.7	
Total capital required	---			---			1,307.0	

	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Cost of coal charge = \$11/Kkg	16,388	65,798	---	---	---	---	---	---
Cost of nuclear fuel = \$0.4929/GJ	---	---	---	---	41,957	---	---	---
Operation and maintenance	---	33,857	---	---	3,887	---	---	---
Subtotal expenses	---	99,655	12.56	---	45,844	5.78	145,499	18.34
Byproduct credits (c)								
Aromatics = \$71/Kkg	1,663	43,082	---	---	---	---	---	---
Ammonia = \$38.5/Kkg NH ₃	253	3,562	---	---	---	---	---	---
Sulfur = \$10/Kkg	393	1,434	---	---	---	---	---	---
Char = /Kkg	2,993	---	---	---	---	---	---	---
Subtotal credits	---	48,078	6.06	---	---	---	48,078	6.06
Expenses after credits	---	---	---	---	---	---	97,421	12.28

TABLE 7-4 (continued)

	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Hydrogen price (utility basis)								
Capital charge (15%)	--	--	--	--	--	--	196,050	24.72
Hydrogen price	--	--	--	--	--	--	--	37.00
Hydrogen price (industry basis)								
Capital charge (25%)	--	--	--	--	--	--	326,750	41.19
Hydrogen price ^(c)	--	--	--	--	--	--	--	53.47

(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

(b) Eight percent per year.

(c) No credit given for methane content of product.

TABLE 7-5
HYDROGEN PRICE ESTIMATE, CASE V, PROCESS TEMPERATURE = 1367°K (2000°F),
HYDROGEN PRODUCED = 28.95 KG/SEC (983 MMSCFD)^(a) AT 0.9 PLANT CAPACITY FACTOR

	Process Plant			HTGR			Total	
Plant investment (\$10 ⁶)								
Process units	561.0			239.0 (NPS)			--	
Off-site facilities	75.0			227.8 (BOP)			--	
Interest during construction ^(b)	105.8			106.7			--	
Total investment	741.8			573.5			1,315.3	
Working capital	--			--			21.6	
Total capital required	--			--			1,336.9	
	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Cost of coal charge = \$11/Kkg	17,124	68,751	7.53	--	--	--	--	--
Cost of nuclear fuel = \$0.4929/GJ	--	--	--	--	41,957	--	--	--
Operation and maintenance	--	35,296	--	--	3,737	--	--	--
Subtotal expenses	--	104,047	11.39	--	45,694	5.01	149,741	16.40
Byproduct credits ^(c)								
Aromatics = \$53/Kkg	1,736	44,977	--	--	--	--	--	--
Ammonia = \$38.5/Kkg NH ₃	266	3,731	--	--	--	--	--	--
Sulfur = \$10/Kkg	411	1,499	--	--	--	--	--	--
Char = /Kkg	3,129	--	--	--	--	--	--	--
Subtotal credits	--	50,207	5.50	--	--	--	50,207	5.50
Expenses after credits	--	--	--	--	--	--	99,534	10.90

TABLE 7-5 (continued)

	Units/cd	\$1000/yr	¢/kg	Units/cd	\$1000/yr	¢/kg	\$1000/yr	¢/kg
Hydrogen price (utility basis)								
Capital charge (15%)	--	--	--	--	--	--	200,535	21.97
Hydrogen price	--	--	--	--	--	--	--	32.87
Hydrogen price (industry basis)								
Capital charge (25%)	--	--	--	--	--	--	334,225	36.61
Hydrogen price ^(c)	--	--	--	--	--	--	--	47.51

(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

(b) Eight percent per year.

(c) No credit given for methane content of product.

TABLE 7-6
ANNUAL OPERATING AND MAINTENANCE COSTS, PROCESS PLANT^(a)
(THOUSANDS OF DOLLARS/YR)

	Process Temperature				
	922°K (1200°F)	1033°K (1400°F)	1144°K (1600°F)	1256°K (1800°F)	1367°K (2000°F)
Plant staffing ^(b)					
Operation	2,629	2,587	2,587	2,587	2,587
Maintenance	6,724	7,817	8,028	8,661	8,940
Subtotal	9,353	10,404	10,615	11,248	11,527
Utilities					
Freshwater (\$0.08/Kkg)	209	227	216	218	215
Electric power (\$0.02/kWh)	--	--	--	--	1,577
Subtotal	209	227	216	218	1,792
Consumables	450	789	858	1,283	1,388
Catalysts and royalties	4,417	6,530	7,057	9,626	8,751
Maintenance ^(b)	6,724	7,817	8,028	8,661	8,940
Miscellaneous ^(c)	2,305	2,620	2,698	2,821	2,898
Total operation and maintenance costs	23,458	28,387	29,472	33,857	35,296

(a) July 1974 basis.

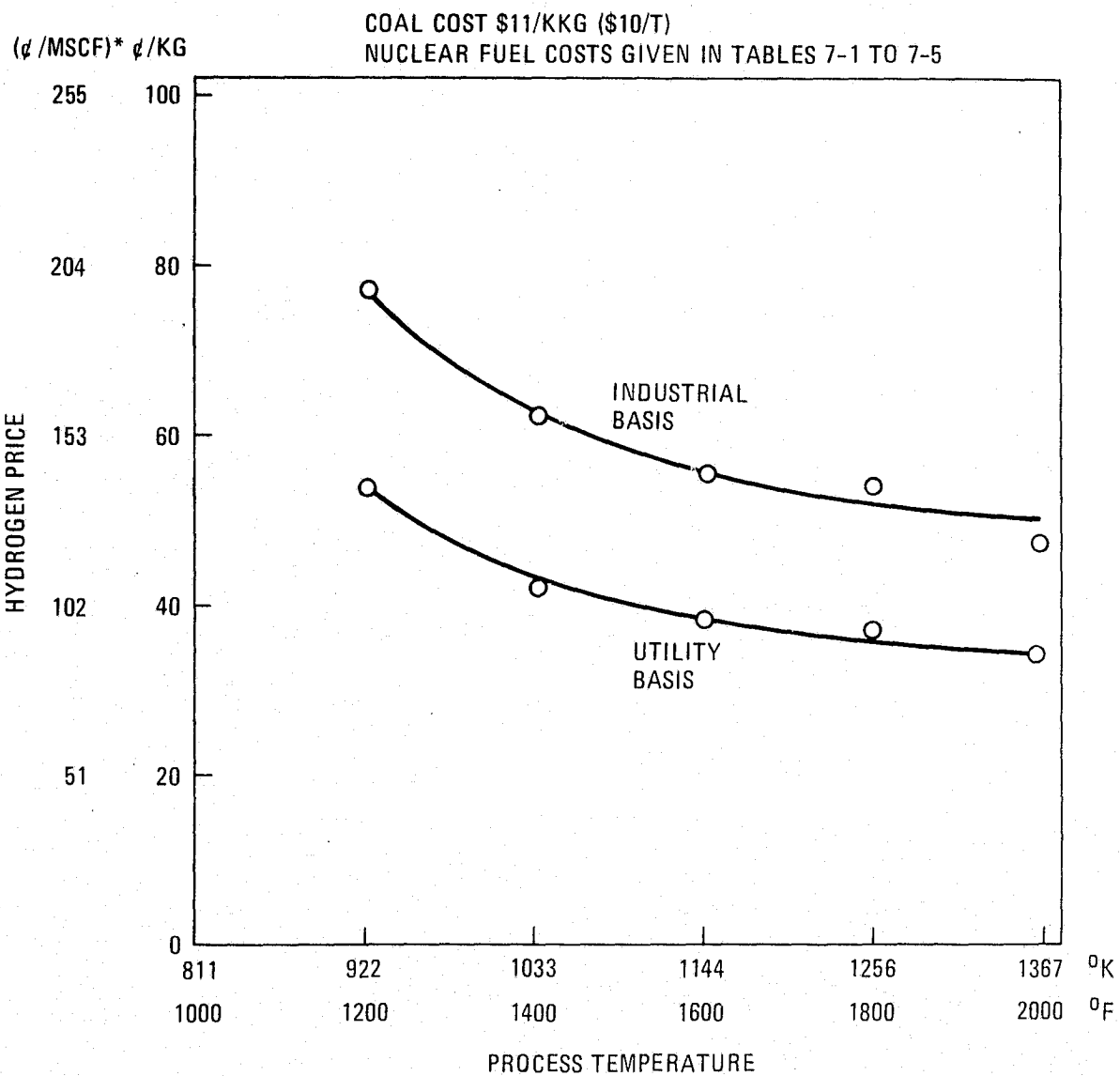
(b) Operating staff charged \$9.50/hr (includes fringe benefits and supervision); maintenance costs calculated to be 3% of capital cost for process units and 1.4% of off-site facility costs divided equally among labor and materials.

(c) Twenty-five percent of total staffing costs.

TABLE 7-7
ANNUAL OPERATION AND MAINTENANCE COSTS, NUCLEAR^(a)
(THOUSANDS OF DOLLARS/YR)

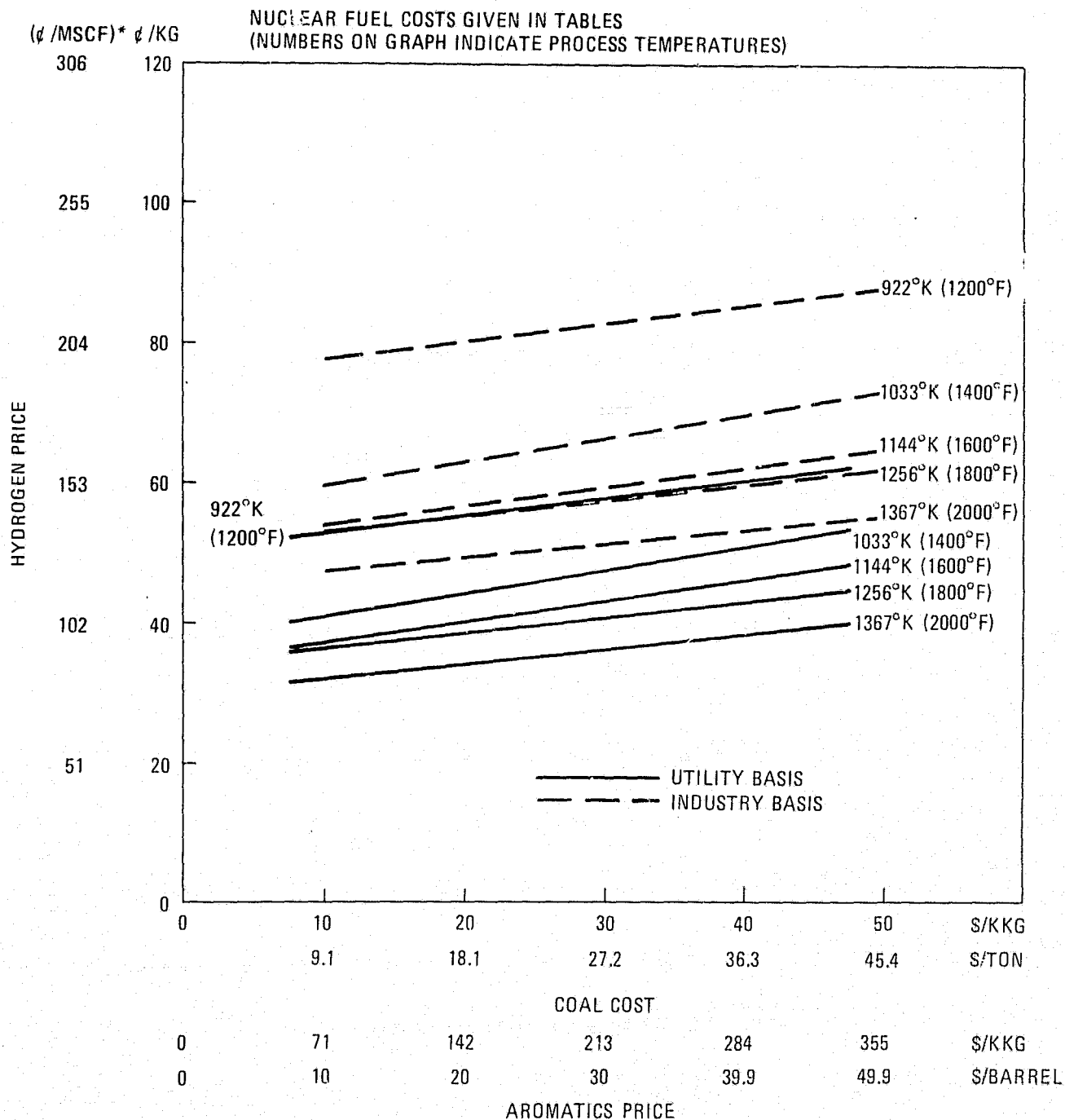
	Process Temperature				
	922°K (1200°F)	1033°K (1400°F)	1144°K (1600°F)	1256°K (1800°F)	1367°K (2000°F)
Station staffing	1640	1640	1737	1814	1814
Contract in-service inspection	20	20	20	20	20
Consumables	486	486	559	559	559
Waste disposal	56	56	58	58	58
Catalyst					
Replacement	846	578	271	230	115
Labor	121	88	48	56	28
Disposal	55	37	18	15	8
Building heating and auxiliary boiler	97	97	97	97	97
Communications	35	35	35	35	35
Maintenance supplies	139	139	139	146	146
Nuclear insurance, commercial	300	300	300	300	300
Nuclear insurance, Price-Anderson	90	90	90	90	90
Annual ERDA license fee	195	195	195	195	195
Miscellaneous operation and maintenance costs	246	246	260	272	272
Total operation and maintenance costs	4326	4007	3827	3887	3737

(a) July 1974 basis.



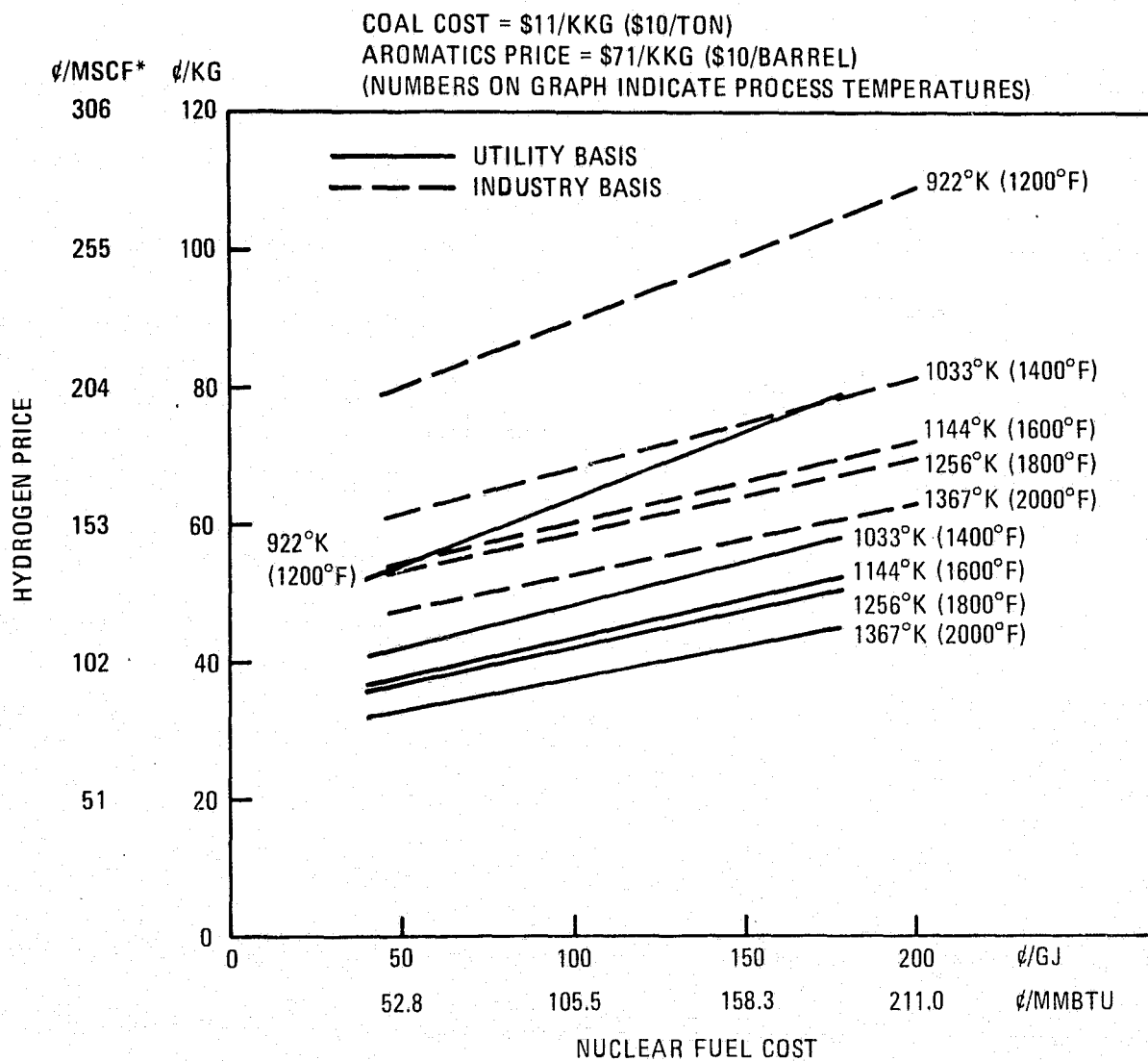
*AT 273°K (32°F), 0.1013 MN/m² (14.7 PSIA).

Fig. 7-1. Hydrogen price (July 1974) as a function of process (reforming) temperature (no credit given for methane content of product)



*AT 273°K (32°F), 0.1013 MN/m² (14.7 PSIA)

Fig. 7-2. Hydrogen price as a function of coal cost (July 1974)
(no credit given for methane content of product)



*AT 273°K (32°F), 0.1013 MN/m² (14.7 PSIA)

Fig. 7-3. Hydrogen price as a function of nuclear fuel cost (July 1974)
 (no credit given for methane content of product)

from \$0.443/GJ to \$0.493/GJ (\$0.467/MMBtu to \$0.521/MMBtu). Nuclear fuel cost has a small but significant effect on hydrogen selling price. Increasing the nuclear fuel cost by a factor of 2 adds \$0.05/kg to \$0.10/kg to the price of hydrogen.

REFERENCE

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8. ENVIRONMENTAL CONSIDERATIONS

8.1. INTRODUCTION

The environmental impact of a hydrogen production unit of the capacity described will be large. In order to produce the approximately 5.0×10^9 kg (5.51×10^6 short tons) of bituminous coal each year for a period of 20 yr, a large surface area has to be mined. The mine by itself will be about 2.5 times larger than the largest mine currently in operation in the U.S. [approximately 2×10^6 short tons/yr (Ref. 8-1)]. It is assumed that the mine and the conversion plant will be located near surface water reservoirs having a capacity sufficient to deliver the required water flows, near railroads, highways, and gas pipelines; the distance between the center of the mine and the coal handling area will not be more than 12 km (8.5 miles).

A comparison of this process for the production of hydrogen with a conventional route (Lurgi) shows considerable advantages for the nuclear route in terms of coal usage and effluent release levels as shown in Section 8.7.

This section is divided into 6 parts: mining, coal handling and storage, coal conversion, product transport, fuel resources, and process comparison.

8.2. MINING

Coal conversion systems are based on either surface or underground mining. Surface, or strip, mining disturbs large amounts of land and often leads to acid mine drainage and silt runoff, both of which degrade water

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quality. A common belief is that adverse environmental impacts of coal mining result only from surface mining; however, underground mining also results in acid drainage and can cause land subsidence over mined-out areas (158,000 acres in Appalachia). Surface mining has been assumed for this study. In general, it can be stated that for economic reasons surface mining is preferred in locations where the ratio of overburden thickness to coal thickness is not greater than 25:1.

The yearly coal requirements are 5.0×10^9 kg (5.51×10^6 short tons). For a seam thickness of 1 - 2m (3 - 6 ft), a surface area of 50 - 100 km² (12,000 - 25,000 acres) for a 20-yr operating period, and 100 - 200 km² (25,000 - 50,000 acres) for a 40-yr period would be required. Seam thicknesses are typical for current mining practice (see Fig. 8-1).

Surface mining is normally done by large equipment drag line, power shovel, or bucket wheel excavators. The coal will be loaded into large trucks by smaller loading shovels for haulage from the mine.

The environmental impact of surface mining depends on the original natural situation as well as the final use after reclamation. Current law in Oklahoma, for example, requires that coal mine operators grade all spoiled ridges and peaks to a rolling topography traversable by machines or equipment customarily utilized in connection with the use to be made of the land after reclamation (Ref. 8-2). The cost of reclaiming the land depends on seam thickness, amount of overburden, and planned recovery time. The shorter the specified recovery time, the more expensive. Rough estimates (Ref. 8-3) showed that an expenditure of approximately $\$7.4 \times 10^5/\text{km}^2$ (\$1.67/metric ton of coal, \$1.51/short ton of coal, or \$0.06/10⁶ Btu) is required for recovery to agricultural land within a period of 50 yr, but for 100% recovery to forest land in 100 yr, the expenditure amounts to $\$2.5 \times 10^5/\text{km}^2$ (\$0.55/metric ton of coal, \$0.50/short ton of coal, \$0.02/10⁶ Btu).

For levels of recovery expenditure which are too low, some types of land may fail to recover at all in reasonable periods of time. For

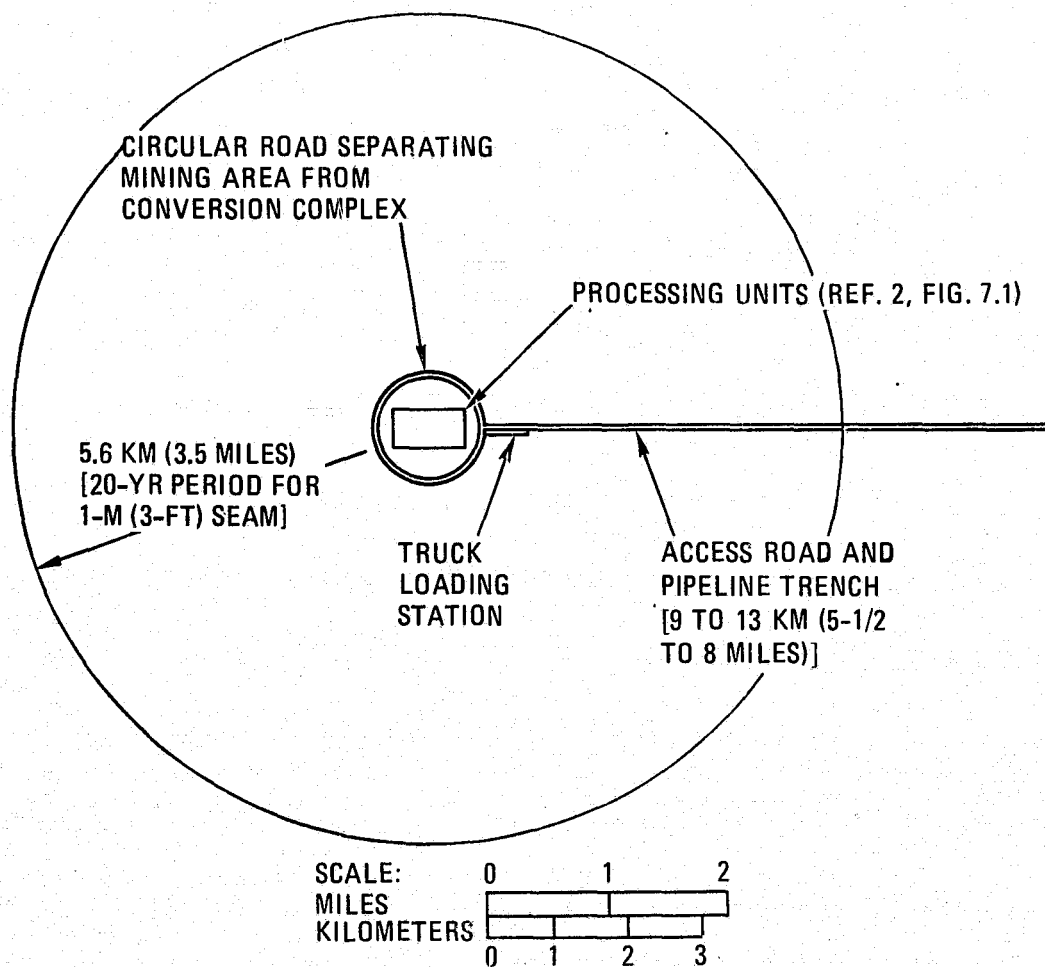


Fig. 8-1. Map showing the mining area and the conversion complex for a 20-yr period

recovery to forest land in a period shorter than 200 yr, the minimum investment is approximately $\$1 \times 10^5/\text{km}^2$ (\$400/acre), and for recovery of agricultural land within 50 yr, a minimum of about $\$5 \times 10^5/\text{km}^2$ (\$2000/acre) has to be spent. Firm data on recovery costs for surface-mined land are sketchy. German experience in the Rhineland brown coal lignite fields suggests a recovery figure of $\$7.5 \times 10^5/\text{km}^2$ (\$3000/acre) to $\$1.1 \times 10^5/\text{km}^2$ (\$440/acre) to obtain agricultural land in about 5 yr. As a first approximation, it is estimated that the total reconstruction cost, assuming optimum control during the stripping operation, will be approximately $\$7$ to $\$9 \times 10^5/\text{km}^2$ (Ref. 8-3), or $\$1.58$ to $\$2.03/\text{metric ton}$ of coal, or $\$0.057$ to $\$0.074/10^6$ Btu.

Mining activity exposes pyritic materials, which become oxidized. Water leaches out acid and metallic salts, which are carried into the local groundwater system and the streams. Because mine drainage is very difficult to control once it occurs, efforts will be made to mine in such a way that no acid drainage will develop. According to Ref. 8-4, the techniques used to avoid mine drainage are based on reclamation of surface lands and waste piles and diversion of water from pollution sources. Reference 8-4 also states that if pollution nevertheless does occur, the polluted water or acid-producing materials can be isolated by deep-well injection and surface and subsurface impoundments; the polluted waters can finally be treated by neutralization processes or diluted by diversion from one water shed to another. The choice of pollution control techniques depends on geologic and hydrologic characteristics of the mine source, the anticipated secondary effects, and the expected water quality improvement. Relatively inexpensive techniques are land reclamation and drainage diversion. Estimates by the U.S. Department of the Interior indicate that the control cost for acid drainage would run about $\$0.10/\text{metric ton}$ ($\$0.09/\text{short ton}$) of coal produced.

8.3. COAL HANDLING AND STORAGE

The environmental impact of coal handling depends on the layout of the mine and the location of the conversion units with respect to the mine.

For transport of the total daily production of $\sim 15 \times 10^6$ kg (16,530 short tons) of coal from the mine to the conversion complex, a combination of 240-ton trucks and conveyor belts will be used. It is anticipated that the coal will be loaded into these large-capacity trucks by loading shovels for haulage from the various mining areas to the coal receiving station. Compared with the environmental impact of the surface mining operation itself, the influence of the transport step between the mining area and the coal receiving station can be neglected. The reconstruction requirements and costs are assumed to be equal to those for the rest of the mine area.

After crushing, the coal will be moved in covered conveyer belts to the pulverized coal area, thus preventing any possible dust problems. Once the coal has been pulverized, it will be pneumatically transported into closed storage bins located inside buildings. The vents of all these operations will be sent to a wet wash tower, and the suspended coal dust in the water will then be sent to a settling pond, where wetting agents will be added to separate the coal from the water. Using current techniques, the coal will be extracted from the settling pond and returned to the mine for disposal. Other methods, such as pelletizing or briquetting and recycling to the coal preparation unit, can be employed when economically feasible.

8.4. COAL CONVERSION

In general, it can be stated that if coal, the most abundant fossil fuel resource of the U.S., is burned in a conventional way without the precautions required by current laws to produce electricity, it causes serious environmental pollution. It produces air pollution in the form of particulates, sulfur oxides, nitrogen oxides, and CO, and it also produces water pollution during mining in the form of siltation and acid mine drainage.

Of course, most of the adverse effects of the use of coal can presently be offset by the use of adequate pollution abatement techniques which aim at reduction of pollutants to allowable levels. The coal

conversion process discussed is basically designed to convert coal feed into clean fuel, in this case hydrogen, and to remove and convert potential pollutants in the processing sequence. When burned, the clean fuel will cause no environmental pollution.

The market study (Section 4) shows that most of the H_2 produced by the present process would not be burned, but would be used for hydrotreating, hydrocracking, hydrodesulfurization, ammonia production, etc. Even for these applications, it is advantageous to upgrade the process gas at one specialized location instead of at numerous places all over the country. It is felt that the environmental impact of this large conversion plant, equipped with highly sophisticated units to prevent environmental pollution, is to be preferred above a great number of small conversion units spread all over the country. This section discusses only the environmental impact of the process portion of the plant. For nuclear environmental considerations, see Refs. 8-2 and 8-5.

Residues from coal gasification processes include solids, liquids, and gases. The solid residue produced in the fluid bed coker is ash which generally consists of about 50% carbon. The material is dry and has a low hydrogen content, and consequently a high ignition temperature. It is estimated (Ref. 8-2) that the percentage of sulfur in the solid residue will be equal to the percentage of sulfur in the coal feed. The solid refuse from the fluid bed coker will be transferred to the storage bins, from where it will be returned to the mine after being sprayed with a conditioner to prevent dusting. The quantity of ash which must be returned to the mine equals ~ 35 kg/sec (~ 3300 short tons/day), i.e., about one-sixth of the total daily consumption of coal. It is assumed that the same 240-ton trucks used for coal transport will be used. Further investigations have to be made to determine the leachable components of the solid refuse and the possible environmental consequences.

Liquid waste in coal gasification consists of water contaminated with phenols, heavy tars, oil, and particulates. Water will be produced in the coal drying operation [~ 10 kg/sec (~ 160 gpm)] and during conversion of the

oxygen in the coal to water [~ 8 kg/sec (~ 130 gpm)]. The water stream is condensed from the gas feed to the hydrogasifier while ammonia is simultaneously condensed and absorbed by the water and the bulk of the H_2S . H_2S which does not react with the ammoniacal liquor is removed at the purification step following the hydrogasifier step. The condensate will be stripped of ammonia, which will be concentrated for sale while the H_2S is sent to the Claus unit for conversion to elemental sulfur. The condensate, containing no suspended solids, will eventually, if phenols are present, be bacteriologically treated to reduce the phenols to acceptable levels, be clarified, and be returned to the cooling water tower system to provide the makeup. It is assumed that most of the phenols will decompose in the hydrogasifier.

The amount of heavy tars produced in the process depends on the operating conditions. It is assumed that the heavy tars can be mixed with the light aromatics when they are used as fuel. Alternately, the heavy tars can be recycled to the fluid bed coker for conversion into coke and distillate fuels.

The flue gas from the fluid bed coker and the H_2S from the acid gas removal system are fed to the sulfur production area. For each 100 kg (220 lb) of sulfur in the coal feed, about 20 kg (44 lb) will be discharged with the ash, and approximately 80 kg (176 lb) will be converted to H_2S , ultimately leaving the plant as solid sulfur after the H_2S has been converted in a conventional Claus oven system. To remove the remaining 4% to 5% of sulfur present in the Claus off-gas, the Claus unit is operated in conjunction with a SCOT plant, in which the SO_2 content of the stack gas is reduced to about 250 ppm. The total sulfur recovery efficiency of a Claus unit followed by a SCOT unit is 99.8% of sulfur intake. By using these highly sophisticated sulfur recovery plants, only ~ 600 tons of SO_2 will be emitted each year. Since less than 1% of the coal carbon is burned in the fluid bed coker to produce flue gas for the Claus unit, no problems are anticipated in venting nitrogen oxides resulting from combustion to the atmosphere.

The proposed plant design is based on the assumption that thermal pollution can be sufficiently limited by the use of cooling towers. If a more detailed analysis proves that this assumption is not justified, thermal pollution effects can be further reduced by air cooling. A large amount of the water required can be eliminated by the use of dry cooling, but the impact on economics has not been explored.

Although several potentially hazardous trace pollutants in coal are present only in the ppm range, it should be noted that the present coal gasification plant handles 173 kg/sec, or 15,000 metric tons (~16,500 short tons), of coal per day. For a potential pollutant, even in the 1 ppm concentration range, a realistic estimate of the quantities might be 6 to 18×10^{-5} kg/sec (10 to 30 lb/day). Since no detailed coal analysis is available, the results of Ref. 8-6 have been used to estimate which quantities of various elements can be expected in the coal feed (see Table 8-1). Part of the metals entering the processing units will leave the units in the ash, and another part will be disposed of on the surface of the catalyst of the hydrocracking unit. Possible adverse effects of the elements on the ecological system and means to avoid this impact will have to be determined by a more elaborate study.

8.5. PRODUCT TRANSPORT

Three well developed techniques exist for transporting the product of the coal conversion plant: by pipeline as a gas, by trailers or rail transport tank cars as liquid hydrogen [$T = 20.4^\circ\text{K}$ (-422.7°F) at 10^5 N/m^2 (14.5 psia)], and by trailers or rail transport tank cars as slush hydrogen [$T = 13.8^\circ\text{K}$ (-452.6°F) at 10^5 N/m^2 (14.5 psia)].

8.5.1. Transport by Pipeline

Hydrogen can be transmitted from the generating station to the consumer areas in underground pipelines similar to those used for natural gas. Numerous refinery and chemical plant hydrogen distribution grids are

TABLE 8-1
ESTIMATE OF ELEMENT QUANTITIES IN COAL FEED FOR GASIFICATION
PLANT [YEARLY COAL INTAKE = 5×10^9 KG (5.51×10^6 SHORT TONS)]

<u>Element</u>	<u>Yearly Quantity</u> <u>$\times 10^3$ kg ($\times 10^3$ lb)</u>
Antimony	7 (15.4)
Arsenic	64 (141.1)
Bromine	68 (149.9)
Cadmium	4 (8.8)
Chromium	77 (169.8)
Cobalt	27 (59.5)
Copper	68 (149.9)
Fluorine	450 (992.2)
Gallium	16 (35.3)
Lead	45 (99.2)
Manganese	245 (540.2)
Mercury	1 (2.2)
Nickel	70 (154.3)
Selenium	10 (22.0)
Vanadium	110 (242.5)
Total	1262 (2782.7)

in use all over the world for short distances. Pipelines carrying hydrogen over distances of up to 80 km (50 miles) are in operation in Texas, and in the Ruhr area of Germany, a hydrogen delivery pipeline network extends over more than 200 km (125 miles). This network has been operative since 1940. The design requirements for gas transmission systems are well understood, and the practices and codes have been well developed. If the technique can be improved to seal existing redundant natural gas mains and piping, these networks can be used to distribute the hydrogen product gas. Experience has shown that the hydrogen transport costs are comparable to those for natural gas, i.e., roughly 10% of the value transported, and the energy loss per 1000 km (~600 miles) due to pumping is about 1%. The ratio of the capital costs of H_2 to CH_4 as well as the ratio of the transportation costs of H_2 to CH_4 equal ~1.44 and ~1.36, respectively. From an environmental point of view, the big difference between hydrogen and natural gas, or methane, is the fire hazard. The explosive range for methane goes from a lower limit of 5.3% to an upper limit of 14.0%, and the range for hydrogen extends from 4.1% to 74.2%. Furthermore, the diffusivity and buoyancy of hydrogen make it very difficult to locate the exact site of a leak. However, excellent safety procedures have been developed for hydrogen in industry and aerospace.

8.5.2. Transportation As Liquid or Slush Hydrogen

Storage and distribution of liquid hydrogen from the point of manufacture to use sites are also well established techniques. Superinsulated 29.5-m^3 (7800-gal) liquid hydrogen trailers and 107-m^3 (28,300-gal) rail transport cars are currently in use, with evaporation losses of no more than $\pm 2\%$ for a 4-day trip. Extensive experience, especially in the NASA space program, has shown that liquid hydrogen can be safely and economically transported across the U.S. in insulated tank cars. Transport of the product as liquid or slush hydrogen will only be used for sites without a gas distribution grid and locations where total consumption is too low to make an extension of the grid attractive. Assuming that an average of 5% of daily production will have to be transported as a liquid,

100 semitrailers will be needed daily. In comparison, a takeoff of the Saturn V rocket for an Apollo mission to the moon required 92 tank trucks, each with a 25-m^3 (6600-gal) capacity, during the last 32 hr. Since this operation has been acknowledged as quite an achievement in safe logistics (Ref. 8-7), it can be stated that the assumed transport of only 5% of daily production by railroad car or truck does constitute an environmental risk, especially to the neighborhood directly next to the conversion plant. Nevertheless, considering its properties, liquid hydrogen, when handled carefully, obviously would not give a worse performance from a safety point of view. It is the scale of the operation which increases the problem. If such an accident happens, the environmental impact will be restricted to the neighborhood directly next to the place of the accident. It has been assumed in the foregoing that all the H_2 produced can be absorbed by the "traditional" consumers. If the H_2 produced has to be used for a hydrogen economy, the environmental problems will be different.

8.6. FUEL RESOURCES

The conversion plant described is designed to produce hydrogen from coal using nuclear heat. As shown in the process description of the Lurgi plant (process 4), hydrogen can also be produced in a classical coal gasification plant, whereby part of the coal is burned to generate the heat for the endothermic reactions. According to the assessments made, $1.06\text{ kg coal/normal m}^3\text{ gas}$ ($0.063\text{ lb coal/scf gas}$) is needed for the Lurgi route, and $0.60\text{ kg coal/normal m}^3\text{ gas}$ ($0.035\text{ lb coal/scf gas}$) (43% less) is needed if the required heat is generated by a process heat HTGR. Thus, U.S. coal reserves can be greatly preserved, and CO_2 , SO_2 , and particulate emissions and production costs of hydrogen can eventually be reduced, depending on the coal price. This means a higher load on uranium resources. Estimates of uranium resources in the U.S. (Ref. 8-8) have shown that the total known quantity of recoverable U_3O_8 is equal to $765 \times 10^6\text{ kg}$ (857×10^3 short tons). Additional resources are estimated to be $1.5 \times 10^9\text{ kg}$ (1700×10^3 short tons) of U_3O_8 . In both cases, a maximum recovery cost of $\$33/\text{kg}$ ($\$15/\text{lb}$) for U_3O_8 has been assumed.

The data indicate that, based on maximum forward costs of \$33/kg (\$15/lb) of U_3O_8 over a 20-yr operating period, 800 plants of the size of the present nuclear plant will consume all known reserves, according to Ref. 8-7, or 200 plants according to Ref. 8-8. Although new reserves with forward costs of up to \$60/kg (\$27/lb) are known and exist in large quantities, these figures nevertheless show that eventually a breeder reactor with an adequately high breeding ratio has to be commercialized in order to cope with the growth rate for electrical as well as process heat capacity.

Besides an average yearly consumption of 64×10^3 kg (141×10^3 lb) of U_3O_8 per year [which is equal to approximately 0.3×10^8 kg (0.66×10^8 lb) crude ore per year and 400 kg (882 lb) of U-235], the HTGR also needs an average yearly input of approximately 9×10^3 kg (9.9 short tons) of ThO_2 , which must be mined and processed up to the required purity. Thorium is essentially produced as a byproduct of the monazite operations for rare earth elements. An excessive stock of thorium-bearing raw materials has therefore always been available. According to Ref. 8-1, 1,789 Mg of ThO_2 equivalent were held in a government stockpile because there was no response to the request for bids during 1972. Hence, for the time being, no new thorium mines have to be set up, and consequently no additional environmental problems are expected. However, if the number of HTGR plants increases, the amount of byproduct thorium will not suffice, and new thorium mines will have to be set up in order to supply the required nuclear fuel.

The environmental impact of fuel processing and fuel reprocessing is not discussed in this section because it is partly covered in Refs. 8-2 and 8-5. A more elaborate discussion is outside the scope of this technological assessment.

8.7. PROCESS COMPARISON

A comparison has been made between the environmental impact of the present nuclear-heated process and the currently commercial hydrogen-producing process utilizing coal for heating, i.e., the Lurgi coal gasification process. This comparison is based on the assumption that both plants produce the same quantity of H_2 , i.e., $269 \text{ m}^3/\text{sec}$ of approximately the same purity, that all pollution abatement equipment is of the same design and efficiency for both the Lurgi process and the nuclear process, and that the energy required for the fresh nuclear fuel and the fuel recycle is generated in a power plant which is also provided with pollution abatement equipment. The pollution caused by the auxiliary units is also taken into account. The environmental impact data have been listed in Table 8-2. It is important to note that all the data in this table are based on calculations and assumptions for specific sites. They have been generated as an illustration of the differences that can be expected. One of the most striking advantages of the nuclear route is a reduction by 43% of the required quantity of coal which, of course, also means a substantial reduction (by a factor of ~ 2.3) in surface mined land and in all adverse effects of surface mining mentioned in Section 8.2. Moreover, the nuclear route implies a significant stretching of the limited coal resources. Since approximately 27% of the coal input of the Lurgi gasifiers leaves the plant as solid waste, the total solid waste production of the Lurgi route is ~ 2.3 times higher than that of the nuclear route.

As regards process water usage, the data show that the water required for the nuclear route, including the nuclear fuel production, is approximately 78% of that required for the Lurgi route.

The big difference between the NO_x emissions of the two processes is caused by the fact that only 1% of the coal intake is burned in the fluid bed coker of the nuclear process, whereas 39.6% of the coal intake of the Lurgi process (equal to 113 kg/sec) is burned in the Lurgi route. No NO_x controls have been incorporated in the two plants.

TABLE 8-2
ENVIRONMENTAL IMPACT OF HYDROGEN PRODUCTION USING NUCLEAR-HEATED REFORMING AND LURGI GASIFICATION OF COAL

Process	Feedstock		Land Use (km ²) (20 yr) (a)	Solid Waste [kg (x 10 ⁶)] (Annual)	Consumed Process Water Use [kg (x 10 ⁶)] (Annual)	SO _x [kg (x 10 ⁶)] (Annual)	NO _x [kg (x 10 ⁵)] (Annual)	C _n H _m , Co [kg (x 10 ³)] (Annual)	Cooling Water Use [kg (x 10 ⁶)] (Annual)
Lurgi	Coal (8.1 x 10 ⁹ kg/yr)	Surface mining	80-160	550	212	--	--	--	
		Processing	8.0	682	10,000	725	22,860	1,600	47,700 ^(b)
		Total	88-168	1,232	10,212	725	22,860	1,600	
Nuclear	Coal (4.9 x 10 ⁹ kg/yr)	Surface mining	50-100	300	118	--	--	--	
		Processing	8.0	972 ^(c)	7,130	275	373	--	24,000 ^(b)
	Nuclear fuel (29.2 x 10 ⁶ kg U ₃ O ₈ /yr)	Mining	5	--	155	3	1.7	0.1	
		Processing	2.5 ^(d)	30 ^(e)	401 ^(f)	73 ^(g)	390	14	14,000 ^(b)
		Total	65.5-115.5	1,302	7,804	351	764.7	14.1	38,000 ^(b)

(a) One-meter and two-meter seam thickness.

(b) Could be lowered significantly by using air cooling.

(c) Includes char; could be significantly lowered by further char utilization.

(d) Permanently committed for tailing ponds 0.1 km², temporary committed ~2.4 km².

(e) Tailings from milling operation.

(f) Water requirements for the fuel cycle plant only. Includes cooling and processing.

(g) Assuming 95% stack gas desulfurization of the power plant generating the required power for enrichment.

It can be concluded that the major advantage of the nuclear route results from a reduction of the coal requirements and a significantly lower NO_x emission.

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9. TECHNOLOGY ASSESSMENT

A technology assessment of the production of hydrogen from coal by hydrogasification is given in Section 3.1 (process 2, hydrogen from coal). This section discusses the effects of process temperature (reforming temperature) on the overall process. These effects are principally confined to the hydrogen production and the hydrogen purification processing units.

9.1. PLANT PERFORMANCE

A summary of plant performance for the five process temperatures selected is given in Table 9-1. The hydrogen produced for a given HTGR thermal capacity increases with increasing process temperature over the entire temperature range studied and is most sensitive in the low temperature range [922° to 1033°K (1200° to 1400°F)]. In this low temperature range, hydrogen production is limited by the fraction of HTGR heat available at temperatures sufficiently high to drive the reforming reaction. Below 1033°K (1400°F), a significant fraction of the HTGR thermal power is devoted to the production of electric power.

The process efficiency is shown as a function of process temperature in Fig. 9-1. Process efficiencies were calculated from the gross heating value of the process inputs and products. Char and acid gas effluents were considered to be waste heat. The efficiency increases with increasing process temperature, closely paralleling the hydrogen production rate. Above process temperatures of 1033°K (1400°F), the production of hydrogen from coal is accomplished with relatively high efficiency. The strong dependence of process efficiency on process temperature in the low temperature range is partly a result of the low efficiency associated with the production of electric power as a byproduct.

TABLE 9-1
SUMMARY OF PLANT PERFORMANCE
(100% CAPACITY)

	Case I	Case II	Case III	Case IV	Case V
Process temperature (°K)	922	1033	1144	1256	1367
(°F)	1200	1400	1600	1800	2000
Coal feed (kg/sec)	98.66	186.71	195.66	210.75	220.21
(tpd)	9396	17,782	18,634	20,071	20,972
HTGR thermal power (MW)	3000	3000	3000	3000	3000
Hydrogen produced (kg/sec)	15.32	22.47	25.99	27.95	32.16
(MMscfd) ^(a)	520	763	882	949	1092
Hydrogen purity (%)	95.4	87.9	90.8	91.0	93.8
Byproduct aromatics (kg/sec)	10.19	18.93	19.75	21.38	22.32
bpd	5794	10,763	11,230	12,156	12,691
Excess power (electric) (MW)	138	57	31	18	7
Plant efficiency	52.4	62.7	65.9	66.9	68.8

^(a) At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

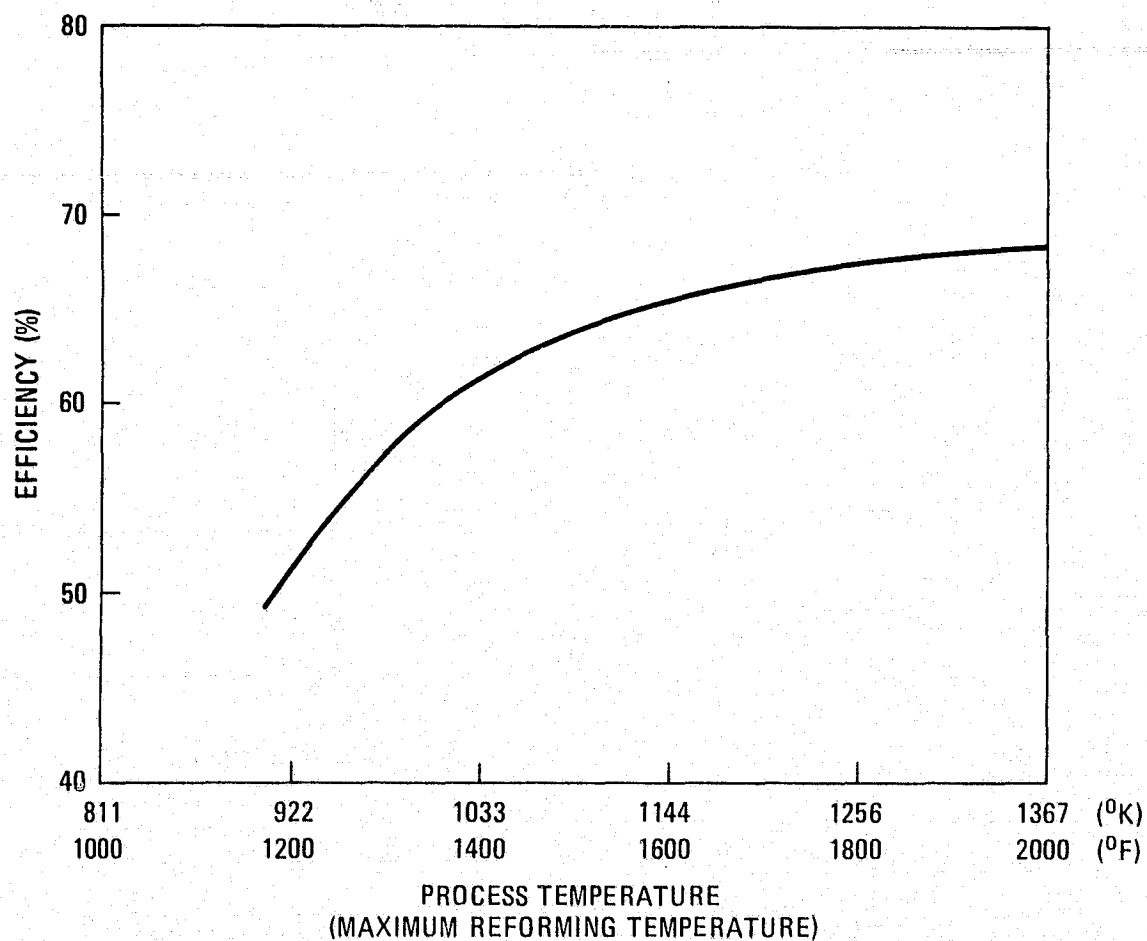


Fig. 9-1. Process efficiency as a function of process temperature

Assessing the relative merits of hydrogen production versus combined hydrogen and electric power production is difficult at this stage of development. However, for production of hydrogen, process efficiency provides a strong incentive for process temperatures of 1033°K (1400°F) or higher. The benefits of process temperatures above 1144°K (1600°F) are small in this study.

9.2. DEVELOPMENT STATUS

The status of process and materials development for the base case [process temperature = 1033°K (1400°F)] is discussed in detail below. The process is technically feasible, but further development work is required to determine economic feasibility. For the other cases in the process study, only the effects of process temperature on the development status of the base case plant are discussed.

9.2.1. Case II, Process Temperature = 1033°K (1400°F)

9.2.1.1. Coal Solubilization. Gulf Research and Development Corporation is currently developing a catalytic coal liquids process. Conversion of over 90% of the coal to liquids has been demonstrated in a 3-lb/day pilot plant. Start-up of a 3-ton/day pilot plant is under way.

Noncatalytic coal solubilization has been demonstrated at the bench scale in several studies. Consolidation Coal Company has successfully operated a 20-ton/day pilot plant.

In summary, solubilization of coals to produce distillable liquids is feasible. Further development work is required to establish optimum operating conditions.

9.2.1.2. Solution Hydrocracking. Hydrocracking processes have been directed toward the production of liquid products. Consequently, temperatures have been limited to between 700° to 720°K (800° to 850°F),

and highly active catalysts have been required. The catalytic process is reasonably well developed in petroleum refining.

For ultimate production of a gaseous product, higher operating temperatures may be desirable. Further experimental work will be required to establish suitable operating conditions for hydrocracking of coal liquids and assessment of the merits of hydrocracking at higher temperatures.

9.2.1.3. Fluid Bed Coking. Fluid bed coking is currently carried out in petroleum refineries using vacuum tower bottoms as feedstock. This material is probably very similar to the coker feed in this process. However, normal coker feedstock contains only small quantities of ash, and the principal area of uncertainty is ash distribution. Alternate solutions to the problem of ash separation exist, including partial oxidation and filtration.

9.2.1.4. Hydrogasification. A limited amount of data exists on thermal hydrogenation of aromatic and paraffinic liquids. One area of uncertainty is the extent of coke formation from thermal cracking and the associated operating conditions required to eliminate coke formation while maintaining a high conversion of hydrogen. However, it appears that distillate feedstocks can be gasified with low production of net carbon. Hydrogasification of petroleum crudes has been demonstrated in commercial units by the British Gas Council.

Experimental work is in progress on hydrogasification of coal liquids. Initial experiments at low pressures are encouraging. Further experimental work is required to establish suitable operating conditions and provide data needed for the design of hydrogasification reactors.

9.2.1.5. Hydrogen Production and Power Generation. The HTGR for electric power production has been under development at GA since 1957. The development effort includes the 40-MW(e) prototype plant at Peach Bottom

(Pennsylvania), which was successfully operated from 1968 to 1974. The 300-MW(e) HTGR at Fort St. Vrain (Colorado) is scheduled for commercial operation in 1975. Over the past 12 yr, a substantial research and development effort has been directed toward the large, 3000-MW(t) and 2000-MW(t) HTGRs.

The use of the HTGR for production of hydrogen by steam-methane reforming requires design modifications in several areas and a limited amount of new development work. However, the basic technology of the HTGR remains applicable to a great extent. The status of the present technology and associated programs required for the development of the process heat HTGR have been discussed in detail in previous studies (Ref. 9-1). A summary is given below.

The fuel developed for the HTGR electric power plants remains applicable. Continuing programs to improve HTGR fuel for electric power production would also yield improvements in process heat reactors.

Materials and components in the nuclear core of HTGR electric power plants are capable of service at the higher temperatures considered in this report. Some modifications to current designs would be required.

Temperature and pressure conditions are within the limitations of current steam generator designs. The development effort would be limited to design modifications.

Design modifications of the helium circulators would be required to handle higher helium temperatures, higher power requirements, and high-pressure steam in the turbine drive. Full-scale testing would be required.

Fossil-fired reformers have been successfully operated on a commercial scale for many years. The use of convectively heated reformers represents a significant departure from conventional practice. However, initial studies have indicated that convectively heated reformers are feasible with

reaction conditions similar to those used in conventional practice. Results of full-scale single-tube tests conducted by Kernforschungsanlage in Germany have further demonstrated the feasibility of nuclear-heated reforming.

Bench-scale catalyst testing and single-tube reformer tests would be necessary to obtain the data required for reformer design. Conventional high-temperature materials appear to be suitable for use in convectively heated reformers. Qualification of materials for service in nuclear reactors would be required.

The current thermal barrier designs for HTGR electric power plants are not suitable for service at the high helium temperatures in the core outlet duct. Alternate materials capable of service at higher temperatures will be required. Several candidate materials are available. Design studies and experimental testing would be required.

9.2.1.6. Hydrogen Purification and Compression. The purification processing steps closely follow conventional practice in modern hydrogen plants. Because CO_2 removal is a major cost item and a major consumer of process heat and power, additional process studies exploring alternate methods for CO_2 removal are justified. No new development work would be required.

Hydrogen compression can be accomplished with conventional equipment. However, compression of the large quantities of low-molecular-weight gas required in the process could justify the development of new compression equipment better suited for the task.

9.2.2. Case I, Process Temperature = 922°K (1200°F)

The development status of coal processing from solubilization through hydrogasification is the same as that for case II [process temperature = 1033°K (1400°F)]. The status of the development of the HTGR and components

is similar to the base case. The helium temperature levels for this case are similar to the helium temperatures in HTGR electric power plants. Core modifications would be minimal and present thermal barrier designs would be adequate. Reformer tube temperatures allow easier solutions to several problems associated with the reformer design. However, catalyst testing and single-tube reformer tests would still be required.

9.2.3. Case III, Process Temperature = 1144°K (1600°F)

The development status of the coal processing units from coal solubilization through hydrogasification remains unchanged. Higher helium and process temperatures require a significant materials development effort for the reformer and the PCRV thermal barrier.

The status of HTGR development for process temperatures of 1144°K (1600°F) is discussed in Ref. 9-1, and general conclusions are summarized below.

Fuel developed for the HTGR electric power plants remains applicable.

The iron base alloys currently used for the cladding and support spine for the control rods in HTGR electric power plants are unsuitable for service at the higher temperatures in the nuclear core. Alternate candidate materials are available, including boronated graphite compacts, coated refractory metals, and several nonmetallic materials. Additional core development work would be limited to design modifications.

A development effort for steam generators would be limited to design modifications [see case II, process temperature = 1033°K (1400°F)].

A development effort for helium circulators would be limited to design modifications [see case II, process temperature = 1033°K (1400°F)].

The high reformer temperatures in this case dictate the use of advanced metallic or nonmetallic materials for the reformer tubes. An extensive materials development program emphasizing ceramics would be required. Extensive studies on reformer design with these materials would be an integral part of the program. As in case II, catalyst testing and full-scale reformer tube testing would be required.

Problem areas for the PCRV are confined to the thermal barrier and are similar to those for the base case [case II, process temperature = 1033°K (1400°F)].

9.2.4. Case IV, Process Temperature = 1256°K (1800°F)

The status of the development of the coal processing units from coal solubilization through hydrogasification remains unchanged. Although the Purisol process was used for hydrogen purification instead of the more conventional hot carbonate process, the Purisol process has been successfully demonstrated in commercial units. Higher helium and process temperatures have a substantial impact on the status of HTGR development. The results of previous studies on the HTGR (Ref. 9-1) are summarized below.

The high helium temperatures required in this case are beyond the capabilities of current HTGR fuel and core designs. An extensive fuel development program or an extensive redesign of the nuclear core would be required.

The inlet helium temperatures to the steam generator are beyond the present capabilities of HTGR steam generators. Design modifications would be required.

Design modifications to the helium circulators would be required [see case II, process temperature = 1033°K (1400°F)].

High temperatures dictate the use of ceramic materials for reformer tubing [see case III, process temperature = 1144°K (1600°F)].

As in the base case, the major problems of the PCRVR are confined to the thermal barrier. However, the problems are more severe in this case owing to higher operating temperatures. The development effort would closely follow that for the base case but would be broader in scope.

9.2.5. Case V, Process Temperature = 1367°K (2000°F)

The development status of the coal processing units from coal solubilization through hydrogasification remains unchanged. The development status of the HTGR is identical to case IV [process temperature = 1256°K (1800°F)].

9.3. POTENTIAL PROCESS IMPROVEMENTS

9.3.1. Review of Process Study

The production of hydrogen was studied for reformer temperatures ranging from 922°K (1200°F) to 1367°K (2000°F). Sufficient time was not available for a detailed study of other process variables. Other major variables closely associated with reformer temperature are process gas pressure and steam-carbon mole ratio in the reformer feed. These variables were intuitively selected based on process and design considerations. Although the results of the process study probably give a reasonably accurate assessment of the benefits of increased temperature, further work in some areas is justified.

In the process study, process heat reactor designs from a previous study (Ref. 9-1) were used as a basis. Overall reformer size was a prime consideration in selecting the process conditions and establishing reformer designs for the process heat reactor. The reformer itself is a major cost item, and reformer size is a major factor in determining the size and cost of the PCRVR and containment structures.

The distribution of HTGR thermal power between the reformer and the steam generators was also a major consideration in selecting process conditions in the reformer. For the low-temperature cases [1033°K (1400°F) and below], hydrogen production is limited by the fraction of HTGR thermal power at temperatures sufficiently high to drive the reforming reaction. The selection of reformer pressure and steam-carbon mole ratio represents a compromise between the desire for high hydrogen purity and small compact reformers. For case I, [process temperature = 922°K (1200°F)], it is possible that lower reformer pressures and increased steam-carbon mole ratios could lead to improvements in the overall process. The more favorable chemical equilibrium under these conditions could allow a greater fraction of HTGR thermal power to be delivered to the reformer. The major uncertainty is the low-temperature activity of the catalyst.

For the higher-temperature cases [1144°K (1600°F) and above], hydrogen production is limited by process heat and power required for purification, compression, and generation of reformer steam. The process gas pressure was increased and the steam carbon mole ratio decreased with increases in process temperature to minimize process heat and power requirements. For the two highest temperatures [1256°K (1800°F) and 1367°K (2000°F)], the Purisol process for hydrogen purification was used to further reduce process heat requirements. Further improvements in the process may be possible for these cases. At process temperatures between 1033° and 1144°K (1400° and 1600°F), the use of an organic solvent CO₂ removal process (such as the Purisol process) could lead to significant increases in hydrogen production; this possibility should be investigated. For case IV [process temperature = 1256°K (1800°F)] and case V [process temperature = 1367°K (2000°F)], the benefits of low steam-carbon mole ratios are questionable because additional steam was required for the shift reaction. Higher steam-carbon mole ratios in the reformer feed and higher conversions are alternate possibilities.

9.3.2. Coal Liquefaction and Hydrogasification

Although additional work on coal liquefaction and hydrogasification was not included in the scope of this study, some potential process improvements in these areas should be noted.

A major area of concern in the process is the separation of ash from coal liquids. Approximately 10% of the gross heating value of the coal is discharged as refuse with the ash. With minor modifications, the coking system could be used to convert a larger fraction of the coal carbon to steam or low-Btu gas. In the latter case, steam would be added to the fluid bed combustor air, and H_2S would be scrubbed from the low-Btu gas effluent. Whether such modifications can be justified on an economic basis depends largely on the cost of coal.

The combination of solution hydrocracking and hydrogasification could lead to a more efficient process and reduce capital cost. In the present study, solubilization of coal is performed in the absence of a catalyst and followed by catalytic hydrocracking and hydrogasification. The catalytic coal liquids process currently being developed by Gulf Research and Development Corporation has demonstrated high yields of coal liquids in a single step.

In the present study, essentially all the coal is converted to hydrocarbon gases to produce feed for the reformers. Reforming higher-molecular-weight hydrocarbons is currently being investigated as an alternate approach. The steam-hydrocarbon reforming process has been demonstrated on hydrocarbons as heavy as naptha. Further process improvements could lead to the successful reforming of higher-molecular-weight hydrocarbons. A major area of uncertainty associated with the reforming of coal liquids is the behavior of reforming catalysts exposed to highly aromatic hydrocarbon feeds.

9.4. SUMMARY

The production of hydrogen by hydrogasification of coal and subsequent steam hydrocarbon reforming is technically feasible, with further development work required to determine economic feasibility. Major areas requiring a significant development effort are nuclear-heated reforming and hydrogasification of coal liquids.

The process efficiency provides a strong incentive for process (reforming) temperatures of at least 1033°K (1400°F). Significant improvements in process efficiency could be realized by increasing the process temperature to 1144°K (1600°F). Above 1144°K (1600°F), increases in process efficiency with increasing temperature are marginal.

The development status of the process is strongly affected by process temperature. For process temperatures of 1033°K (1400°F) and below, a limited amount of new development work is required. For process temperatures of 1144°K (1600°F) and above, an extensive development effort is required, concentrating on high-temperature materials for the reformer and the PCRV thermal barrier.

Future efforts should concentrate on process temperatures of about 922°K (1400°F). This temperature level allows for efficient production of hydrogen and a development program with high chances of success. The development program required for commercialization of the process is discussed in Section 10. At process temperatures of 1144°K (1600°F) and above, the development of a whole spectrum of new technology in high-temperature materials is required. The chances of successful commercialization of the process are unknown, and commercialization efforts should be considered of secondary importance at this stage of development.

REFERENCE

- 9-1. "High-Temperature Nuclear Heat Source Study," USAEC Report GA-A13158, General Atomic, December 31, 1974.

10. COMMERCIALIZATION PROGRAM

The research and development requirements for commercialization of the Stone & Webster/Gulf coal solution gasification process is discussed in this section. The following program has been developed for hydrogen production with a process (reforming) temperature of 1033°K (1400°F). Selection of this temperature allows for efficient production of hydrogen, with a limited development effort required for the HTGR and steam hydrocarbon reformer.

10.1. HTGR

The research and development work required for the first process heat reactor is described below. A more detailed description of the research and development program is given in Ref. 10-1.

10.1.1. Schedule

The first portion of the preliminary design for the process heat HTGR will be devoted to analyzing cycle conditions, identifying component design requirements, and evaluating systems and safety problems. By the end of the preliminary design phase, preliminary component designs should be in hand and should reflect trade-offs between maximizing system performance, minimizing component design problems, and minimizing development program time and expense. Upon completion of the preliminary design, the system should be defined well enough to support detail design.

A major concern in plant design will be the required performance of the reformer and the characteristics of the catalyst. In the program schedule shown in Fig. 10-1, initial testing on the reformer begins after

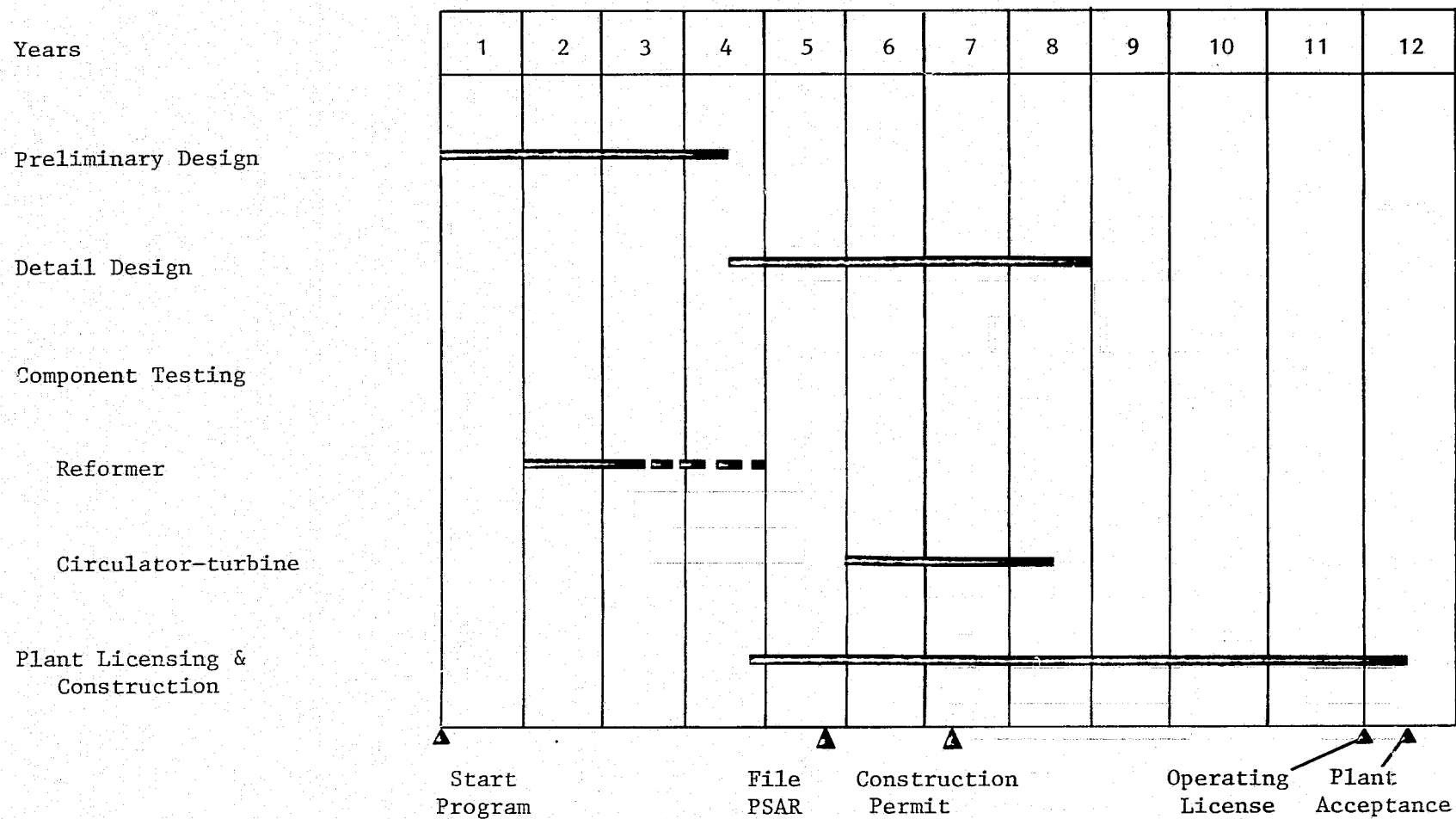


Fig. 10-1. Program schedule for hydrocarbon reforming

1/2 yr of preliminary design and extends for a total of 1-1/2 yr. Initially, the test work will be used to obtain basic performance data and catalyst characteristics in support of reformer design studies. The test program will evolve to a demonstration and proof test on a full-scale reformer tube(s). This work is considered to be part of the research and development program.

The licensing and construction segment is more time-consuming than any other phase of the program. Because of the unique combination of nuclear reactor and reformer, the construction licensing process is expected to be longer than that for conventional HTGR power plants.

10.1.2. Significant Components

Major development work centers about three basic changes to conventional HTGR designs. The major change is the addition of a helium-heated steam-methane reformer in a wall cavity of the PCRV. The reformer itself represents the one new piece of equipment in the HTGR. Although this reformer is similar in size to the existing steam generators and can be installed and removed much as they are, it is expected that the design will significantly differ from that of the steam generators. However, the experience and technical information gained in developing the HTGR steam generator will be most valuable.

The second major change is the helium temperature. The principal research effort in this area will be devoted to selection and testing of new or modified PCRV thermal barrier. The presence of helium temperatures above the standard steam plant temperatures will also require a limited amount of additional work on core design modifications. Differences between temperature levels have been taken into account in the research and development program.

The elimination of the reheat portion of the steam generator is significant. The main steam conditions are nearly the same as those for

the electrical HTGR plants, but the steam turbine used to drive helium circulator experiences much different steam conditions. The inlet pressure is increased from 6.2 to 17.2 MN/m³ (900 to 2500 psia) and the temperature is increased from 644°K to 786°K (700° to 955°F). In both instances, the turbines would be single-stage, large-flow machines driving nearly identical helium circulators.

The safety implications of coupling a nuclear reactor with a chemical processing plant requires careful technical analysis. The operational requirements and limitations of the two plants must be identified, and a suitable control and safety system must be developed. It is expected that reformer catalyst will have to be replaced periodically. Plant design and safety evaluations will be greatly influenced by this requirement.

10.1.3. Preliminary Design Phase

10.1.3.1. Optimization. The cycle conditions and component designs which constitute the current reference plant were chosen as the most promising among a number of alternatives. It is appropriate to re-examine alternatives in greater depth as more detailed information, namely information generated by more refined component and system design work and systems analysis, becomes available. Real cycle losses, computed component efficiencies, capital and operating cost, etc., may significantly influence the choice of the cycle conditions and basic design features of the plant.

10.1.3.2. Component and Systems Design. The reformer design is the most critical component because it has no direct counterpart in current nuclear plants. It will markedly influence the overall plant configuration. The catalyst used in the reformer has, in turn, the most critical influence on reformer design.

Reformer design work must begin with fundamental examination of chemical and transport phenomena important to catalytic reactions. A major question to be resolved is whether widely used catalysts and reformer

designs are adaptable to nuclear reactors or whether better designs can be developed. Reformer design will be supported by and will closely parallel reformer test programs under the research and development program. Initially, these tests would be fundamental bench-scale tests of catalyst materials and geometries, influence of flow, temperature, and chemical constituents on conversion, etc. Of course, the reformer design would also have to consider structural, mechanical, and thermal performance.

The circulator drive turbine must be sized according to allowable stresses. Its bearings, seals, structural support, and auxiliary systems must be designed and analyzed, and its performance at off-design conditions determined. Adjustments for increased core outlet temperature must be made, and core and fuel design and fuel management must be specified. Insulation and thermal barrier protecting the reactor vessel must be examined for possible adverse conditions. Major characteristics of BOP systems must be determined, including piping sizes, pressure losses, electrical power requirements, water system designs, plant buildings and structures, etc.

10.1.3.3. Systems Analysis. Systems analysis will provide information regarding full-load, off-design, and transient performance and will support plant safety studies and licensing. Such studies will provide insights into the characteristics of such a process heat plant as well as help establish design requirements of plant systems and components. It is expected that computer codes developed for the steam cycle HTGR power plant can be used as the basis for a methods development effort suitable for the HTGR/reformer.

The chemical process has a large potential influence on system characteristics, the most obvious being in the area of licensing and safety, where accidents could be very different from those postulated for an electrical HTGR plant. Hydrogen diffusion through the reformer tube walls has been identified as a potential problem and will be examined as part of the reformer test program. It is important to establish safe and

licensable design guidelines as soon as possible so that plant design can proceed in the most expeditious manner. As analysis continues and design details are more fully identified, the development program will be appropriately updated.

10.1.3.4. Evaluation. Preliminary design work will provide the basis for initiating economic evaluation of the nuclear hydrocarbon reformer concept. Plant costs will be estimated, and preliminary cost/benefit or profitability analyses will be performed based on projected market conditions for such process heat reactors.

10.1.4. Detail Design Phase

The objective of the detail design phase is to completely refine and specify plant design preparatory to construction of the first commercial HTGR/reformer. In addition to critical components whose design and development will have been well under way, specifications and/or designs for the following plant systems will have to be developed:

1. Circulator and circulator turbine auxiliaries.
2. Helium purification system.
3. Plant protection and control systems, especially as required by the reformer.
4. Helium and steam system instrumentation.
5. Feedwater/steam piping, pumps, valves, etc.
6. Cooling water systems.
7. Plant buildings and structures.

8. Electrical system.
9. Water treatment.
10. Steam turbine/electrical generator system.

A Preliminary Safety Analysis Report, a Final Safety Analysis Report, an environmental impact statement, and other assistance which may be required by the plant owner in obtaining licenses will be provided.

10.2. PROCESS PLANT

A four-phase program has been started to achieve the orderly development of the Stone & Webster/Gulf nuclear coal solution gasification process. The phases are as follows:

1. Phase I, bench-scale unit: preparation of special liquid coal feedstocks; design and operation of a bench-scale liquid coal hydrogasification unit; supporting engineering work and conceptual HTGR design work. This work is currently being performed at Gulf Research & Development Co. facilities at Harnarville, Pa.
2. Phase II, pilot plant: design and operation of a larger and hydrogasification plant; engineering work on the process and the HTGR to establish the design basis for a commercial coal conversion project.
3. Phase III, process demonstration plant and HTGR development: if required, a larger-scale facility on the order of 0.2 kg/sec (7 MMscfd)* of gas output to demonstrate mechanical, chemical, and economic features of the process portion of the plant; HTGR

*At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

development work to complete the design basis and the required developmental testing.

4. Phase IV, commercial plant: design and construction of a commercial-size 20.6-kg/sec (700-MMscfd)* (nominal) SPG plant. It is contemplated that the first train of 5 to 7.5 kg/sec (476 to 714 + pd) capacity would be started up with a fossil-fired heat source, with the completed plant utilizing a nuclear heat source.

The entire program has been designed to minimize expenditures during the early stages prior to operation of the demonstration plant and, at the same time, to achieve early completion of phase IV, construction of a commercial plant.

10.3. SUMMARY SCHEDULE

Figure 10-2 shows the summary schedule for the combined nuclear and process portion of the plant. Major tie points on the schedule are (1) the completion of the reference design for the nuclear plant and the start of the demonstration process plant and (2) the completion of construction of the nuclear portion and the remaining trains of the process portion. The total time span is 11 yr.

REFERENCE

- 10-1. "High-Temperature Nuclear Heat Source Study," USAEC Report GA-A13158, General Atomic, December 30, 1974.

*At 273°K (32°F), 0.1013 MN/m² (14.7 psia).

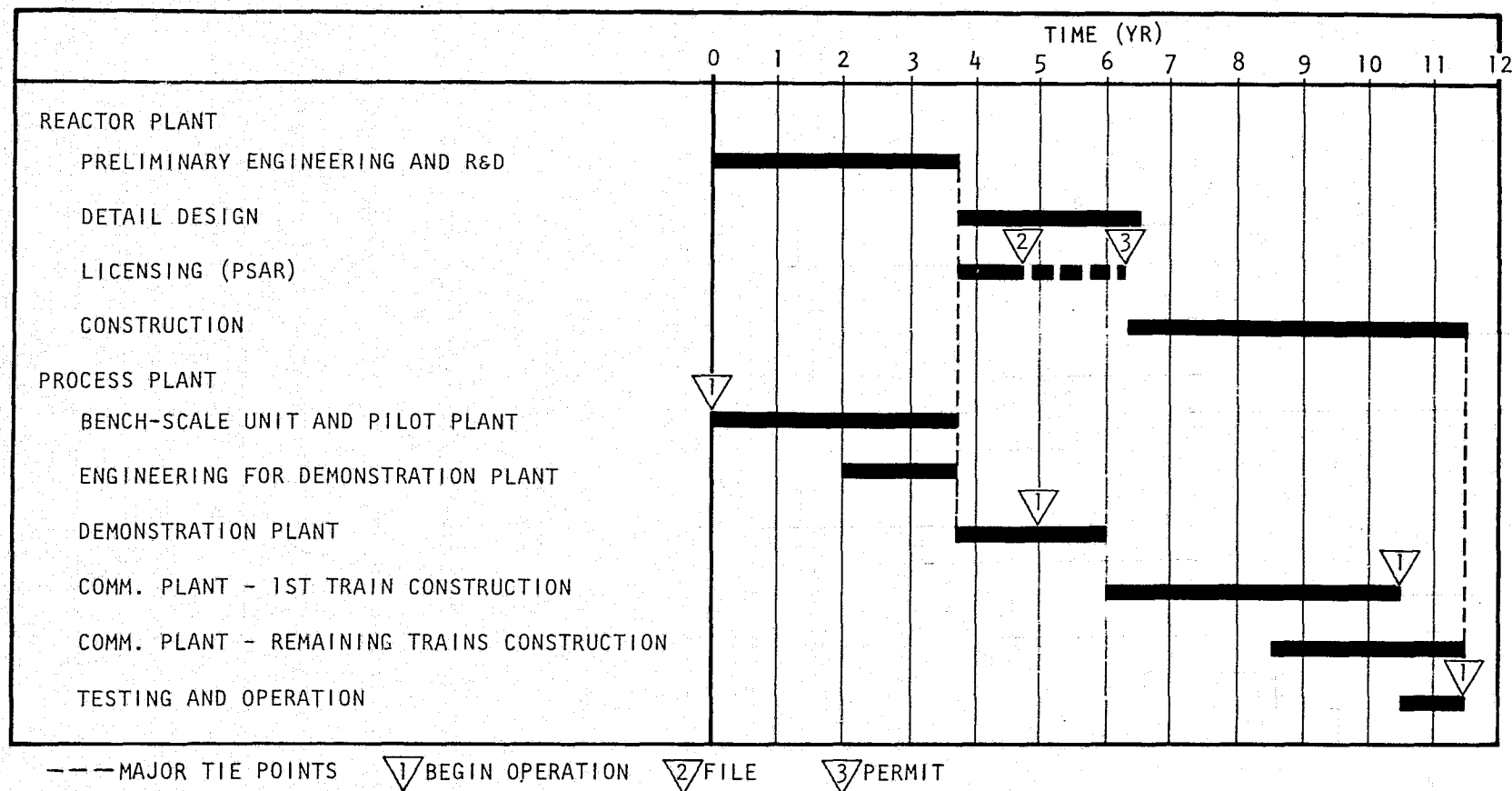


Fig. 10-2. Summary schedule for commercial plant

11. CONCLUSIONS AND RECOMMENDATIONS

The following conclusions have been reached:

1. Several processes for hydrogen production from a variety of feedstocks (light hydrocarbons, coal, water) are potential candidates for future manufacture of hydrogen.
2. From the processes surveyed, it appears that in general, processes utilizing more abundant raw materials are at a more primitive stage of development. It is possible that many of the processes surveyed will reach commercialization in an appropriate sequence of time as the resources of fossil fuel become increasingly more scarce in future years and the technology for utilizing more abundant resources develops.
3. The use of a nuclear heat source in hydrogen production would significantly contribute to fossil fuel conservation efforts.
4. The development status of the nuclear coal solution gasification process is technically feasible at process temperatures of 1033°K (1400°F) and below. For process temperatures of 1144°K (1600°F) and higher, an extensive development program in high-temperature materials with unknown chances of success is required.
5. Process efficiency increases and cost of producing hydrogen decreases with increasing process temperatures, particularly in the temperature range 922°K (1200°F) to 1089°K (1500°F).
6. The nuclear coal solution gasification process is capable of meeting present environmental standards.

7. A hydrogen market survey has shown that the demand for hydrogen will significantly increase by the year 2000, particularly in the manufacture of synthetic fuels. Although it appears that most of the hydrogen will be captively produced by the consuming plant, development of a merchant hydrogen market is not necessarily excluded.

Recommendations are listed below.

1. The first generation HTGR for hydrogen production should be developed for a process temperature of between 1033° to 1144°K (1400° to 1600°F).
2. Development of coal solution and hydrogasification processing should continue in order to determine the economic feasibility of the nuclear coal solution gasification process.
3. Further process studies on the nuclear coal solution gasification process covering a wider scope of process considerations should be conducted in order to develop a more economical process design.
4. Development efforts on process thermochemical water splitting should continue in order to assess the technical feasibility of water splitting processes and to determine the development effort required for commercialization.

APPENDIX A

WATER SPLITTING DETAILS

The process steps in Fig. 3-13 typically contain more than one unit operation, and some secondary material streams were omitted from Figs. 3-14 through 3-19. The unit operations within the processing steps are shown in Fig. A-1; Table A-1 is keyed to the stream numbers of Fig. A-1.

The kinds of equipment added in Fig. A-1 are pumps, power generation devices, some vapor condensers, and some liquid vaporizers which were not shown in previous figures. Many of the streams shown in Fig. A-1 but not in prior figures connect the heat transfer fluid streams of various unit operations.

The material balance shown in Table A-1 is based on a 23% conversion of total heat input into enthalpy of dissociation and work of compression of water at 300°K (80°F) and $1.013 \times 10^5 \text{ MN/m}^2$ (14.7 psia) into hydrogen at 300°K (80°F) and $2.1 \times 10^5 \text{ MN/m}^2$ (30 psia) and into oxygen at 300°K (80°F), 99% of which is at $5 \times 10^6 \text{ MN/m}^2$ (725 psia) and 1% of which is at $1.013 \times 10^5 \text{ MN/m}^2$ (14.7 psia). The mass flow rate of the H_2 product output under these conditions is 4.7 kg/sec (10.36 lb/sec) of H_2 during process operation. Waste H_2 appearing in the waste gas vent stream, stream 16, is 0.51% of the product hydrogen output. Stream numbers are keyed to the streams shown in Fig. A-1.

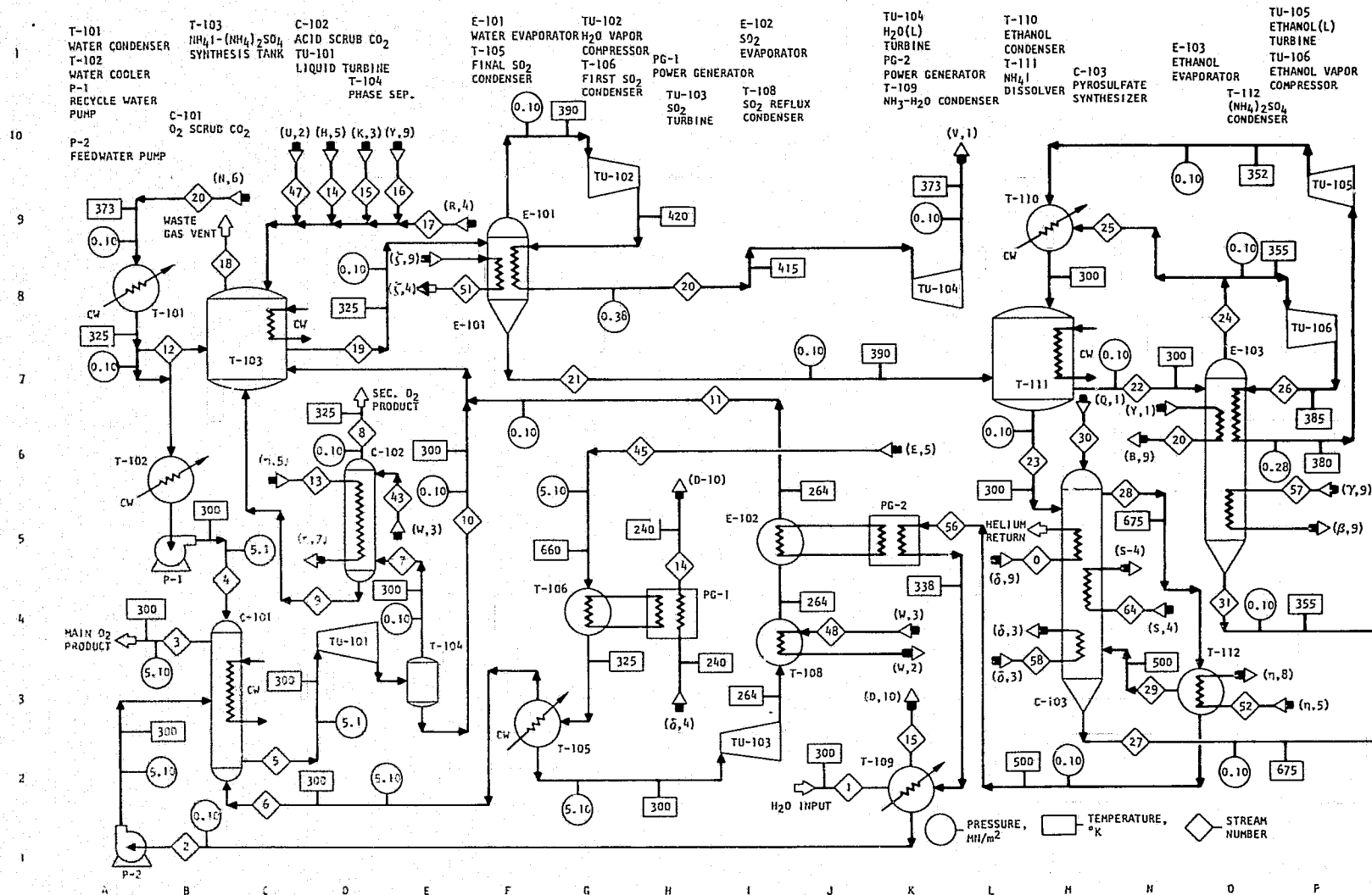


Fig. A-1. Flow sheet for ISPRA MK10 water splitting process (Sheet 1)

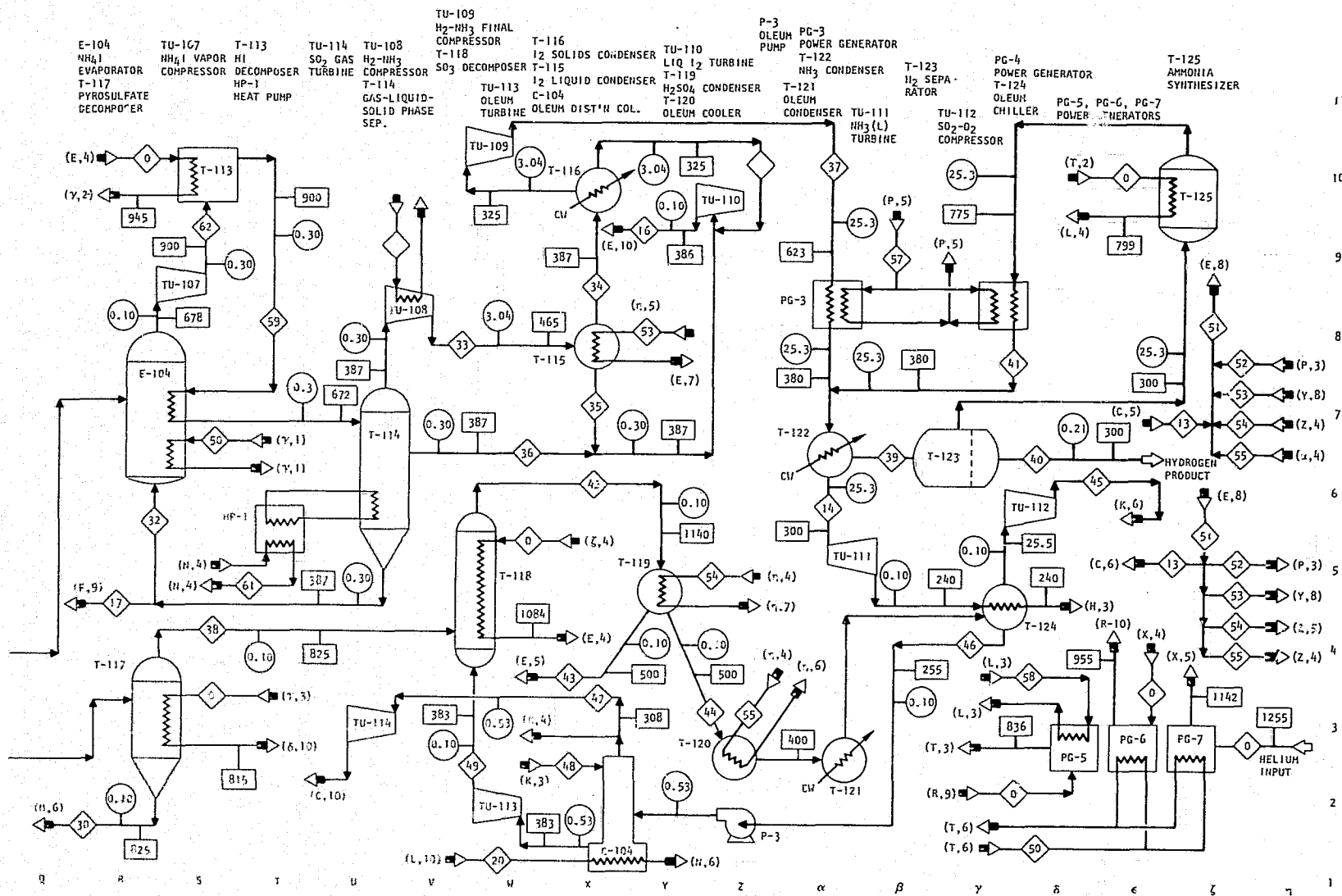


Fig. A-1. Flow sheet for ISPRA MKIO water splitting process (Sheet 2)

TABLE A-1
ISPRA MARK-10 FLOW RATES

Stream		Flow Rate ^(a) (kg-mole/sec)															Total Flow Rate (kg/sec)	
		Total Flow Rate	H ₂	O ₂	I ₂	N ₂	H ₂ O	SO ₂	SO ₃	NH ₃	HI	NH ₄ I	(NH ₄) ₂ SO ₄	Na ₂ SO ₄	Na ₂ S ₂ O ₇	H ₂ SO ₄		Other Constituents
0	Helium	299.026															299.026 helium	1197.0
1	Water input	2.358					2.358											42.48
2	H ₂ O-NH ₃ scrub	5.109					4.550		0.559									91.49
3	Main O ₂ product	1.146		1.137			0.009											36.55
4	H ₂ O scrub	9.325					9.325											168.00
5	Spent scrub	14.240		0.012			13.749		0.326			0.059					0.058(NH ₄) ₂ SO ₃	268.17
6	O ₂ -SO ₂ to scrubbing	1.294		1.177				0.117										45.16
7	Secondary O ₂ -NH ₃	0.013		0.012			T		0.0001									0.386
8	Secondary O ₂ product	0.012		0.012														0.384
9	H ₂ SO ₄ spent scrub	0.157					T					T				0.157		15.40
10	Degassedd spent scrub	14.191		T			13.749		0.325			0.059					0.058(NH ₄) ₂ SO ₃	267.77
11	Recycle SO ₂	2.177		0.023				2.131	0.023									139.09
12	Recycle water	46.627					46.627											840.03
13	Acid scrub col. coolant	(b)																(b)
14	Recycle NH ₃	4.698	0.012		0.0002				4.686									79.89
15	Recycle wet NH ₃	4.837					0.233		4.604									85.87
16	Recycle I ₂ combined	2.345	0.0002		2.343													594.75
17	Recycle NH ₄ I catalyst	0.476			T													69.00
18	Waste off-gas	0.041	0.012	0.023			0.006					0.476						0.868
19	NH ₄ I-(NH ₄) ₂ SO ₄ solution	63.696					55.952					5.162	2.582					2097.5
20	Recycle water, vapor	55.952					55.952											1008.0
21	NH ₄ I-(NH ₄) ₂ SO ₄ salts	7.744					T					5.162	2.582					1089.5
22	NH ₄ I-EtOH solution	51.789										5.162	T				46.627 ethanol	2896.4
23	(NH ₄) ₂ SO ₄ salt	2.582											2.582				Trace ethanol	341.19
24	Total ethanol vapor	46.627															46.627 ethanol	2148.1
25	Low-pressure EtOH vapor	3.194															3.194 ethanol	147.15
26	Compressed EtOH vapor	43.433															43.433 ethanol	2001.0
27	Na ₂ S ₂ O ₇ salt	2.838					T							0.099	2.425		0.314 NaHSO ₄	590.36
28	NH ₃ -H ₂ O-SO ₃ vapor	8.474					2.612		0.187	5.629						0.046		162.41
29	Condensed (NH ₄) ₂ SO ₄	0.233											0.233					30.79
30	Recycle Na ₂ SO ₄	2.681												2.681				380.84
31	NH ₄ I salt	5.162										5.162						748.32

TABLE A-1 (Continued)

Stream	Flow Rate ^(a) (kg-mole/sec)															Total Flow Rate (kg/sec)
	Total Flow Rate	H ₂	O ₂	I ₂	N ₂	H ₂ O	SO ₂	SO ₃	NH ₃	HI	NH ₄ I	(NH ₄) ₂ SO ₄	Na ₂ S ₂ O ₇	H ₂ SO ₄	Other Constituents	
32 Recycle NH ₄ I salt	16.297			T							16.297					2362.5
33 Impure H ₂	7.949	3.253		0.313	0.303				4.080							163.99
34 H ₂ ; low I ₂	7.647	3.252		0.012	0.303				4.080							87.54
35 I ₂ liquid recycle 1	0.301	0.0002		0.301												76.41
36 I ₂ liquid recycle 2	2.033			2.033												516.06
37 H ₂ ; trace I ₂	7.635	3.252		0.0002	0.303				4.080							84.59
38 SO ₃ gas	2.734					0.153		2.577						0.004		209.07
39 H ₂ ; low NH ₃	18.278	16.203			1.515				0.560							84.65
40 Pure H ₂ product	2.331	2.331														4.700
41 NH ₃ -H ₂ recycle	15.340	12.962			1.212				1.166							80.12
42 SO ₃ dissociation products	4.439		1.201			0.273	2.402	0.563								242.30
43 Sulfuric acid	0.157						T	T						0.157		15.40
44 Cooled SO ₃ dissociation products	4.009		1.201				2.402	0.290						0.116		226.90
45 O ₂ -SO ₂ gas	3.471		1.201				2.247	0.023								184.24
46 Oleum-SO ₂ liquid	0.537						0.154	0.267						0.116		42.62
47 SO ₂ still vapor	0.154						0.154									9.86
48 SO ₂ still reflux	0.154						0.154									9.86
49 SO ₂ still bottoms	0.383							0.267						0.116		32.75
50 NH ₄ I dissociation heat fluid	(b)															(b)
51 H ₂ O evaporator heat fluid	(b)															(b)
52 (NH ₄) ₂ SO ₄ condenser coolant	(b)															(b)
53 I ₂ (liquid) condenser coolant	(b)															(b)
54 H ₂ SO ₄ condenser coolant	(b)															(b)
55 Oleum cooler coolant	(b)															(b)
56 NH ₃ -H ₂ O vapor recycle	7.588					2.425			5.163							131.63
57 EtOH evaporator heat fluid	(b)															(b)
58 Pyrosulfate decomposer heat fluid	(b)															(b)
59 NH ₄ I decomposer output	43.524	3.253		2.343	0.303				20.853	16.772						3110.8
60 I ₂ solid recycle	0.012			0.012												3.05
61 Pyrosulfate decomposer heat fluid	(b)															(b)
62 Input decomposer	42.918								21.459	21.459						3110.8

(a) T = trace; 0 < T < 0.005.

(b) Flow rate and composition of heating/cooling fluid may be chosen for convenience.

APPENDIX B

LISTS OF MAJOR EQUIPMENT

The abbreviations listed below are used in the lists contained in this appendix.

S/T = shell/tube,
CS = carbon steel,
SS = stainless steel,
OVHD = overhead,
CW = cold water,
BFW = boiler feedwater,
HTS = high-temperature shift,
LTS = low-temperature shift.

SECTIONS 100/200
COAL PREPARATION/SOLUTION
CASE 1
TEMPERATURE = 922°K (1200°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
G-108 A/D	Pulverizers with feeders and drives	4	23 kg/sec each (91 short tons/hr)				Total drives = 1810 kW (2428 hp)
F-201 A/B	Solvent heater	2	(Fired) 36,000 kW (123 MMBtu/hr)			18-35% Cr 8-20% Ni	Process outlet temperature = 922°K (1200°F)
L-201	Coal liquefaction reactor	1	3.96 m ϕ x 40.85 m (13 ft ϕ x 134 ft)	533 (500)	15.160 (2200)	SA-299 Liner: 2-1/4 Cr - 1 Mo Internal insulation: 20 cm (8 in.)	Hemispherical heads
M-201	Coal slurry drum	1	3.99 m ϕ x 13.72 m (13 ft 1 in. ϕ x 45 ft)	505 (450)	0.345 (50)	SA-299	Conical bottom
M-203	Coal slurry pre-heat drum	1	1.98 m ϕ x 8.08 m (6 ft 6 in. ϕ x 26 ft 6 in.)	533 (500)	15.160 (2200)	SA-535	
M-204	Solvent surge drum	1	2.31 m ϕ x 17.07 m (7 ft 7 in. ϕ x 56 ft)	644 (700)	1.723 (250)	SA-299	
M-205	First flash drum	1	2.31 m ϕ x 9.15 m (7 ft 7 in. ϕ x 30 ft)	708 (815)	3.446 (500)	SA-533	
M-211	Vacuum drum	1	4.93 m ϕ x 27.44 m (16 ft 2 in. ϕ x 90 ft)	366 (200)	Vac (vac)	SA-299	
M-213	Vacuum flash separator	1	2.31 m ϕ x 12.20 m (7 ft 7 in. ϕ x 40 ft)	366 (200)	Vac (vac)	SA-299	
P-201 A/D	Coal slurry circulation pump	4	Total = 0.0948 m ³ /sec (1503 gpm)	477 (400)	$\Delta P = 0.345$ (50)	Centrifugal	Motor drive total = 43 kW (58 hp)
P-204 A/F	Coal liquefaction feed pump	6	Total = 0.0471 m ³ /sec (746 gpm)	477 (400)	$\Delta P = 14.678$ (2130)	Centrifugal	Motor drive total = 808 kW (1084 hp)
P-205 A/B	Solvent circulation pump	2	Total = 0.0689 m ³ /sec (1092 gpm)	589 (600)	$\Delta P = 1.378$ (200)	Centrifugal	Motor drive total = 130 kW (174 hp)
P-209	Third flash product pump	1	Total = 0.0061 m ³ /sec (96 gpm)	589 (600)	$\Delta P = 12.404$ (1800)	Centrifugal	Motor drive total = 377 kW (506 hp)
P-211 A/B	Vacuum drum bottoms pump	2	Total = 0.0293 m ³ /sec (465 gpm)	366 (200)	$\Delta P = 13.093$ (1900)	Centrifugal	Motor drive total = 593 kW (795 hp)
Q-202	Solvent storage tank	1	6893 m ³ (43,350 bbl)	422 (300)		SA-299	Cone roof tank
				S/T	S/T	S/T	
T-211	Vacuum flash trim condenser	1	Total = 5112 m ² (55,000 ft ²)	505/339 (450/150)	0.345/vac/0.861 (50/vac/125)	CS/CS	10 shells
T-212	Vacuum flash condenser	1	Total = 3765 m ² (40,500 ft ²)	686/616 (775/650)	0.345/vac/1.723 (50/vac/250)	CS/CS	8 shells
T-216	Vacuum drum feed heater	1	Total = 2696 m ² (29,000 ft ²)	811/930 (1000/1215)	0.861/1.723 (125/250)	2-1/4 Cr - 1 Mo/304 SS	5 shells

SECTIONS 100/200
COAL PREPARATION/SOLUTION
CASE 2
TEMPERATURE = 1033°K (1400°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
G-108 A/F	Pulverizers with feeders and drives	6	29.0 kg/sec each (115 short tons/hr)				Total drives = 3426 kW (4595 hp)
F-201 A/B	Solvent heater	2	(Fired) 68,270 kW (233 MMBtu/hr)			18-35% Cr 8-20% Ni	Process outlet temperature = 922°K (1200°F)
L-201 A/B	Coal liquefaction reactor	2	3.86 m ϕ x 40.85 m (12 ft 8 in. ϕ x 134 ft)	533 (500)	15.160 (2200)	SA-299 Liner: 2-1/4 Cr - 1 Mo Internal insulation: 20 cm (8 in.)	Hemispherical heads
M-201 A/B	Coal slurry drum	2	3.89 m ϕ x 13.72 m (12 ft 9 in. ϕ x 45 ft)	505 (450)	0.345 (50)	SA-299	Conical bottom
M-203 A/B	Coal slurry pre-heat drum	2	1.96 m ϕ x 8.08 m (6 ft 5 in. ϕ x 26 ft 6 in.)	533 (500)	15.160 (2200)	SA-533	
M-204	Solvent surge drum	1	3.18 m ϕ x 17.07 m (10 ft 5 in. ϕ x 56 ft)	644 (700)	1.723 (250)	SA-299	
M-205	First flash drum	1	3.18 m ϕ x 9.15 m (10 ft 5 in. ϕ x 30 ft)	708 (815)	3.446 (500)	SA-533	
M-211	Vacuum drum	1	6.78 m ϕ x 27.44 m (22 ft 3 in. ϕ x 90 ft)	366 (200)	Vac (vac)	SA-299	
M-213	Vacuum flash separator	1	3.18 m ϕ x 12.20 m (10 ft 5 in. ϕ x 40 ft)	366 (200)	Vac (vac)	SA-299	
P-201 A/D	Coal slurry circulation pump	4	Total = 0.1798 m ³ /sec (2850 gpm)	477 (400)	ΔP = 0.345 (50)	Centrifugal	Motor drive total = 82 kW (110 hp)
P-204 A/F	Coal liquefaction feed pump	6	Total = 0.0891 m ³ /sec (1412 gpm)	477 (400)	ΔP = 14.678 (2130)	Centrifugal	Motor drive total = 1530 kW (2052 hp)
P-205 A/B	Solvent circulation pump	2	Total = 0.1306 m ³ /sec (2070 gpm)	589 (600)	ΔP = 1.378 (200)	Centrifugal	Motor drive total = 246 kW (330 hp)
P-209 A/B	Third Flash product pump	2	Total = 0.0119 m ³ /sec (182 gpm)	589 (600)	ΔP = 12.404 (1800)	Centrifugal	Motor drive total = 716 kW (960 hp)
P-211 A/B	Vacuum drum bottoms pump	2	Total = 0.0556 m ³ /sec (882 gpm)	366 (200)	ΔP = 13.093 (1900)	Centrifugal	Motor drive total = 1122 kW (1505 hp)
Q-202	Solvent storage tank	1	13,050 m ³ (82,100 bbl)	422 (300)		SA-299	Cone roof tank
				S/T	S/T	S/T	
T-211	Vacuum flash trim condenser	1	Total = 2798 m ² (30,100 ft ²)	505/339 (450/150)	0.345/vac/0.861 (50/vac/125)	CS/CS	5 shells
T-212	Vacuum flash condenser	1	Total = 2059 m ² (22,150 ft ²)	686/616 (775/650)	0.345/vac/1.72 (50/vac/250)	CS/CS	4 shells
T-216	Vacuum drum feed heater	1	Total = 1475 m ² (15,870 ft ²)	811/930 (1000/1215)	0.861/1.723 (125/250)	2-1/4 Cr - 1 Mo/304 SS	3 shells

B-3

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SECTIONS 100/200
COAL PREPARATION/SOLUTION
CASE 3
TEMPERATURE = 1144°K (1600°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
G-108 A/F	Pulverizers with feeders and drives	6	30.3 kg/sec each (120 tons/hr)				Total drives = 3590 kW (4815 hp)
F-201 A/B	Solvent heater	2	(Fixed) 71,200 kW (243 MBtu/hr)			18-35% Cr 8-20% Ni	Process outlet temperature = 922°K (1200°F)
L-201 A/B	Coal liquefaction reactor	2	3.96 m ϕ x 40.85 m (13 ft ϕ x 133 ft)	533 (500)	15.160 (2200)	SA-299 Liner: 2-1/4 Cr - 1 Mo Internal insulation: 20 cm (8 in.)	Hemispherical heads
M-201 A/B	Coal slurry drum	2	3.96 m ϕ x 13.72 m (13 ft ϕ x 45 ft)	505 (450)	0.345 (50)	SA-299	Conical bottom
M-203 A/B	Coal slurry pre-heat drum	2	1.98 m ϕ x 8.08 m (6 ft 6 in. ϕ x 26 ft 6 in.)	533 (500)	15.160 (2200)	SA-533	
M-204	Solvent surge drum	1	3.23 m ϕ x 17.07 m (10 ft 7 in. ϕ x 56 ft)	644 (700)	1.723 (250)	SA-299	
M-205	First flash drum	1	3.23 m ϕ x 9.15 m (10 ft 7 in. ϕ x 30 ft)	708 (815)	3.446 (500)	SA-533	
M-211	Vacuum drum	1	6.91 m ϕ x 27.44 m (22 ft 8 in. ϕ x 90 ft)	366 (200)	Vac (vac)	SA-299	
M-213	Vacuum flash separator	1	3.23 m ϕ x 12.20 m (10 ft 7 in. ϕ x 40 ft)	366 (200)	Vac (vac)	SA-299	
P-201 A/D	Coal slurry circulation pump	4	Total = 0.1880 m ³ /sec (2980 gpm)	477 (400)	ΔP = 0.345 (50)	Centrifugal	Motor drive total = 86 kW (115 hp)
P-204 A/F	Coal liquefaction feed pump	6	Total = 0.0933 m ³ /sec (1479 gpm)	477 (400)	ΔP = 14.678 (2130)	Centrifugal	Motor drive total = 1603 kW (2149 hp)
P-205 A/B	Solvent circulation pump	2	Total = 0.1366 m ³ /sec (2166 gpm)	589 (600)	ΔP = 1.378 (200)	Centrifugal	Motor drive total = 257 kW (344 hp)
P-209 A/B	Third flash pre-heat pump	2	Total = 0.020 m ³ /sec (190 gpm)	589 (600)	ΔP = 12.404 (1800)	Centrifugal	Motor drive total = 748 kW (1003 hp)
P-211 A/B	Vacuum drum bottoms pump	2	Total = 0.0582 m ³ /sec (923 gpm)	366 (200)	ΔP = 13.093 (1900)	Centrifugal	Motor drive total = 1175 kW (1576 hp)
Q-202	Solvent storage tank	1	13,590 m ³ (80,500 bbl)	422 (300)		SA-299	Cone roof tank
				S/T	S/T	S/T	
T-211	Vacuum flash trim condenser	1	Total = 2914 m ² (31,350 ft ²)	505/339 (450/150)	0.345/vac/8.61 (50/vac/125)	CS/CS	5 shells
T-212	Vacuum flash condenser	1	Total = 2147 m ² (23,100 ft ²)	686/616 (775/650)	0.345/vac/1.72 (50/vac/250)	CS/CS	4 shells
T-216	Vacuum drum feed heater	1	Total = 1543 m ² (16,600 ft ²)	811/930 (1000/1215)	0.861/1.723 (125/250)	2-1/4 Cr - 1 Mo/304 SS	3 shells

SECTIONS 100/200
COAL PREPARATION/SOLUTION
CASE 4
TEMPERATURE = 1255°K (1800°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
G-108 A/H	Pulverizers with feeders and drives	8	24.7 kg/sec each (98 tons/hr)				Total drives = 3872 kW (5192 hp)
F-201 A/B	Solvent heater	2	(Fired) 77,060 kW (263 MMbtu/hr)			18-35% Cr 8-20 Ni	Process outlet temperature = 922°K (1200°F)
L-201 A/B	Coal liquefaction reactor	2	4.12 m ϕ x 40.85 m (13 ft 6 in. ϕ x 134 ft)	533 (500)	15.160 (2200)	SA-299 Liner: 2-1/4 Cr - 1 Mo Internal insulation: 20 cm (8 in.)	Hemispherical heads
M-201 A/B	Coal slurry drum	2	4.12 m ϕ x 13.72 m (13 ft 6 in. ϕ x 45 ft)	505 (450)	0.345 (50)	SA-299	Conical bottom
M-203 A/B	Coal slurry pre-heat drum	2	2.08 m ϕ x 8.08 m (6 ft 10 in. ϕ x 26 ft 6 in.)	533 (500)	15.160 (2200)	SA-533	
M-204	Solvent surge drum	1	3.35 m ϕ x 17.07 m (11 ft 0 in. ϕ x 56 ft)	644 (700)	1.723 (250)	SA-299	
M-205	First flash drum	1	3.35 m ϕ x 9.15 m (11 ft 0 in. ϕ x 30 ft)	708 (815)	3.446 (500)	SA-533	
M-211	Vacuum drum	1	7.19 m ϕ x 27.44 m (23 ft 7 in. ϕ x 90 ft)	366 (200)	Vac (vac)	SA-299	
M-213	Vacuum flash separator	1	3.35 m ϕ x 12.20 m (11 ft 0 in. ϕ x 40 ft)	366 (200)	Vac (vac)	SA-299	
P-201 A/D	Coal slurry circulation pump	4	Total = 0.2027 m ³ /sec (3214 gpm)	477 (400)	ΔP = 0.345 (50)	Centrifugal	Motor drive total = 92 kW (124 hp)
P-204 A/F	Coal liquefaction feed pump	6	Total = 0.1006 m ³ /sec (1595 gpm)	477 (400)	ΔP = 14.678 (2130)	Centrifugal	Motor drive total = 1729 kW (2318 hp)
P-205 A/B	Solvent circulation pump	2	Total = 0.1474 m ³ /sec (2336 gpm)	589 (600)	ΔP = 1.378 (200)	Centrifugal	Motor drive total = 276 kW (370 hp)
P-209 A/B	Third flash product pump	2	Total = 0.0130 m ³ /sec (206 gpm)	589 (600)	ΔP = 12.404 (1800)	Centrifugal	Motor drive total = 807 kW (1082 hp)
P-211 A/B	Vacuum drum bottoms pump	2	Total = 0.0628 m ³ /sec (995 gpm)	366 (200)	ΔP = 13.093 (1900)	Centrifugal	Motor drive total = 1245 kW (1670 hp)
Q-202	Solvent storage tank	1	14,740 m ³ (92,700 bbl)	422 (300)		SA-299	Cone roof tank
				S/T	S/T	S/T	
T-211	Vacuum flash trim condenser	1	Total = 3160 m ² (34,000 ft ²)	505/339 (450/150)	0.345/vac/0.861 (50/vac/125)	CS/CS	6 shells
T-212	Vacuum flash condenser	1	Total = 2324 m ² (25,000 ft ²)	686/616 (775/650)	0.345/vac/1.72 (50/vac/250)	CS/CS	5 shells
T-216	Vacuum drum feed heater	1	Total = 1673 m ² (18,000 ft ²)	811/930 (1000/1215)	0.861/1.723 (125/250)	2-1/4 Cr - 1 Mo/304 SS	3 shells

SECTIONS 100/200
COAL PREPARATION/SOLUTION
CASE 5
TEMPERATURE = 1366°K (2000°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
G-108 A/H	Pulverizers with feeders and drives	8	25.7 kg/sec each (102 tons/hr)				Total drives = 4040 kW (5418 hp)
F-201 A/B	Solvent heater	2	(Fired) 80,575 kW (275 MMBtu/hr)			18-35% Cr 8-20% Ni	Process outlet temperature = 922°K (1200°F)
L-201 A/B	Coal liquefaction reactor	2	4.22 m ϕ x 40.85 m (13 ft 10 in. ϕ x 134 ft)	533 (500)	15.160 (2200)	SA-299 Liner: 2-1/4 Cr - 1 Mo Internal insulation: 20 cm (8 in.)	Hemispherical heads
M-201 A/B	Coal slurry drum	2	4.22 m ϕ x 13.72 m (13 ft 10 in. ϕ x 45 ft)	505 (450)	0.345 (50)	SA-299	Conical bottom
M-203 A/B	Coal slurry pre-heat drum	2	2.13 m ϕ x 8.08 m (7 ft 0 in. ϕ x 26 ft 6 in.)	533 (500)	15.160 (2200)	SA-533	
M-204	Solvent surge drum	1	3.43 m ϕ x 17.07 m (11 ft 3 in. ϕ x 56 ft)	644 (700)	1.723 (250)	SA-299	
M-205	First flash drum	1	3.43 m ϕ x 9.15 m (11 ft 3 in. ϕ x 30 ft)	708 (815)	3.446 (500)	SA-533	
M-211	Vacuum drum	1	7.34 m ϕ x 27.44 m (24 ft 1 in. ϕ x 90 ft)	366 (200)	Vac (vac)	SA-299	
M-213	Vacuum flash separator	1	3.43 m ϕ x 12.20 m (11 ft 3 in. ϕ x 40 ft)	366 (200)	Vac (vac)	SA-299	
P-201 A/D	Coal slurry circulation pump	4	Total = 0.2116 m ³ /sec (3354 gpm)	477 (400)	ΔP = 0.345 (50)	Centrifugal	Motor drive total = 96 kW (129 hp)
P-204 A/F	Coal liquefaction feed pump	6	Total = 0.1050 m ³ /sec (1664 gpm)	477 (400)	ΔP = 14.678 (2130)	Centrifugal	Motor drive total = 1804 kW (2419 hp)
P-205 A/B	Solvent circulation pump	2	Total = 0.1538 m ³ /sec (2438 gpm)	589 (600)	ΔP = 1.378 (200)	Centrifugal	Motor drive total = 289 kW (387 hp)
P-209 A/B	Third flash product pump	2	Total = 0.0135 m ³ /sec (214 gpm)	589 (600)	ΔP = 12.404 (1800)	Centrifugal	Motor drive total = 842 kW (1129 hp)
P-211 A/B	Vacuum drum bottoms pump	2	Total = 0.0655 m ³ /sec (1039 gpm)	366 (200)	ΔP = 13.093 (1900)	Centrifugal	Motor drive total = 1323 kW (1774 hp)
Q-202	Solvent storage tank	1	15,380 m ³ (96,750 bbl)	422 (300)		SA-299	Cone roof tank
				S/T	S/T	S/T	
T-211	Vacuum flash trim condenser	1	Total = 3300 m ² (35,500 ft ²)	505/339 (450/150)	0.345/vac/0.861 (50/vac/125)	CS/CS	6 shells
T-212	Vacuum flash condenser	1	Total = 2417 m ² (26,000 ft ²)	686/616 (775/650)	0.345/vac/1.72 (50/vac/250)	CS/CS	5 shells
T-216	Vacuum drum feed heater	1	Total = 1738 m ² (18,700 ft ²)	811/930 (1000/1215)	0.861/1.723 (125/250)	2-1/4 Cr - 1 Mo/304 SS	4 shells

SECTION 300
HYDROCRACKING/SEPARATION
CASE 1
TEMPERATURE = 922°K (1200°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
L-301 A/D	Hydrocracking reactors	4	2.64 m ϕ x 27.44 m (8 ft 8 in. ϕ x 90 ft)	(Metal) 533 (500) 811 (1000)	13.78 (2000)	SA-533 Internal insulation: 20 cm (8 in.)	
				S/T	S/T	S/T	
T-301 A/B	Reactor coolers	2	Total = 558 m ² (6000 ft ²)	394/811 (250/1000)	1.72/13.78 (250/2000)	CS/2-1/4 Cr - 1 Mo	
M-350	High-pressure flash drum	1	2.31 m ϕ x 12.20 m (7 ft 7 in. ϕ x 40 ft)	727 (850)	3.45 (500)	SA-533	
M-353 A/B	Atmospheric flash drum and stripper	2	2.80 m ϕ x 10.98 m (9 ft 2 in. ϕ x 36 ft)	727 (850)	0.345 (50)	SA-533	Includes 4 valve trays each
M-355 A/B	Vacuum flash drum	2	4.93 m ϕ x 28.96 m (16 ft 2 in. ϕ x 95 ft)	727 (850)	0.1/vac (15/vac)	SA-533	
P-352 A/B	Atmospheric flash OVHD product pump	2	Total = 0.017 m ³ /sec (260 gpm)	727 (850)	13.78 (2000)	Centrifugal	Motor drive total = 431 kW (578 hp)
R-350	Flashed vapor booster compressors	1	1.299 kg/sec (10,300 lb/hr) 862 kW (1156 bhp) MW = 8.7			2-stage centrifugal	Suction = 2.75 MN/m ² , 329°K (400 psig, 132°F) Discharge = 11.37 MN/m ² (1650 psig)
T-352	Atmospheric flash OVHD vapor/product exchanger	1	Total = 372 m ² (4000 ft ²)	705/616 (810/650)	0.345/14.47 (50/2100)	CS/CS	2 shells
T-353	Atmospheric flash CW condenser	1	Total = 534 m ² (5750 ft ²)	508/366 (455/200)	0.345/0.517 (50/75)	CS/CS	2 shells
T-354	Vacuum flash feed heater	1	Total = 753 m ² (8100 ft ²)	839/727 (1050/850)	2.067/0.345 (300/50)	CS/CS	2 shells
T-355	Vacuum flash OVHD vapor/product exchanger	1	Total = 1766 m ² (19,000 ft ²)	727/616 (850/650)	0.1/vac/0.86 (15/vac/125)	CS/CS	4 shells
T-356	Vacuum flash CW condenser	1	Total = 1301 m ² (14,000 ft ²)	505/366 (450/200)	0.1/vac/0.52 (15/vac/75)	CS/CS	6 shells

SECTION 300
HYDROCRACKING/SEPARATION
CASE 2
TEMPERATURE = 1033°K (1400°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
L-301 A/D	Hydrocracking reactors	4	3.61 m ϕ x 27.44 m (11 ft 10 in. ϕ x 90 ft)	(Metal) 532 (500) 811 (1000)	13.78 (2000)	SA-533 Internal insulation: 20 cm (8 in.)	
				S/T	S/T	S/T	
T-301 A/B	Reactor coolers	2	Total = 1070 m ² (11,500 ft ²)	394/811 (250/1000)	1.72/13.78 (250/2000)	CS/2-1/4 Cr - 1 Mo	
M-350	High-pressure flash drum	1	3.18 m ϕ x 12.20 m (10 ft 5 in. ϕ x 40 ft)	727 (850)	3.45 (500)	SA-533	
M-353 A/B	Atmospheric flash drum and stripper	2	3.84 m ϕ x 10.98 m (12 ft 7 in. ϕ x 36 ft)	727 (850)	0.345 (50)	SA-533	Includes 4 valve trays each
M-355 A/B	Vacuum flash drum	2	6.78 m ϕ x 28.96 m (22 ft 3 in. ϕ x 95 ft)	727 (850)	0.1/vac (15/vac)	SA-533	
P-352 A/B	Atmospheric flash OVHD product pump	2	Total = 0.031 m ³ /sec (495 gpm)	727 (850)	13.78 (2000)	Centrifugal	Motor drive total = 817 kW (1095 hp)
R-350	Flashed vapor booster compressors	1	2.457 kg/sec (19,500 lb/hr) 1633 kW (2190 bhp) MW = 8.7			2-stage centrifugal	Suction = 2.75 MN/m ² , 329°K (400 psig, 132°F) Discharge = 11.37 MN/m ² (1650 psig)
T-352	Atmospheric flash OVHD vapor/product exchanger	1	Total = 710 m ² (7660 ft ²)	705/616 (810/650)	0.345/14.47 (50/2100)	CS/CS	2 shells
T-353	Atmospheric flash CW condenser	1	Total = 1020 m ² (10,950 ft ²)	508/366 (455/200)	0.345/0.517 (50/75)	CS/CS	2 shells
T-354	Vacuum flash feed heater	1	Total = 1425 m ² (15,320 ft ²)	839/727 (1050/850)	2.067/0.345 (300/50)	CS/CS	2 shells
T-355	Vacuum flash OVHD vapor/product exchanger	1	Total = 3360 m ² (36,100 ft ²)	727/616 (850/650)	0.1/vac/0.86 (15/vac/125)	CS/CS	4 shells
T-356	Vacuum flash CW condenser	1	Total = 2440 m ² (26,260 ft ²)	505/366 (450/200)	0.1/vac/0.52 (15/vac/75)	CS/CS	6 shells

SECTION 300
HYDROCRACKING/SEPARATION
CASE 3
TEMPERATURE = 1144°K (1600°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
L-301 A/D	Hydrocracking reactors	4	3.66 m ϕ x 27.44 m (12 ft ϕ x 90 ft)	(Metal) 533 (500) 811 (1000)	13.78 (2000)	SA-533 Internal insulation: 20 cm (8 in.)	
				S/T	S/T	S/T	
T-301 A/B	Reactor coolers	2	Total = 1115 m ² (12,000 ft ²)	394/811 (250/1000)	1.72/13.78 (250/2000)	CS/2-1/4 Cr - 1 Mo	
M-350	High-pressure flash drum	1	3.23 m ϕ x 12.20 m (10 ft 7 in. ϕ x 40 ft)	727 (850)	3.45 (500)	SA-533	
M-353 A/B	Atmospheric flash drum and stripper	2	3.94 m ϕ x 10.98 m (12 ft 11 in. ϕ x 36 ft)	727 (850)	0.345 (50)	SA-533	Includes 4 valve trays each
M-355 A/B	Vacuum flash drum	2	6.91 m ϕ x 28.96 m (22 ft 8 in. ϕ x 95 ft)	727 (850)	0.1/vac (15/vac)	SA-533	
P-352 A/B	Atmospheric flash OVHD product pump	2	Total = 0.033 m ³ /sec (516 gpm)	727 (850)	13.78 (2000)	Centrifugal	Motor drive total = 855 kW (1146 hp)
R-350	Flashed vapor booster compressors	1	2.570 kg/sec (20,400 lb/hr) 1709 kW (2292 bhp) MW = 8.7			2-stage centrifugal	Suction = 2.75 MN/m ² , 329°K (400 psig, 132°F) Discharge = 11.37 MN/m ² (1650 psig)
T-352	Atmospheric flash OVHD vapor/product exchanger	1	Total = 744 m ² (8000 ft ²)	705/616 (810/650)	0.345/14.47 (50/2100)	CS/CS	4 shells
T-353	Atmospheric flash CW condenser	1	Total = 1069 m ² (11,500 ft ²)	508/366 (455/200)	0.345/0.517 (50/75)	CS/CS	4 shells
T-354	Vacuum flash feed heater	1	Total = 1487 m ² (16,000 ft ²)	839/727 (1050/850)	2.067/0.345 (300/50)	CS/CS	4 shells
T-355	Vacuum flash OVHD vapor/product exchanger	1	Total = 3504 m ² (37,700 ft ²)	727/616 (850/650)	0.1/vac/0.86 (15/vac/125)	CS/CS	7 shells
T-356	Vacuum flash CW condenser	1	Total = 2556 m ² (27,500 ft ²)	505/366 (450/200)	0.1/vac/0.52 (15/vac/75)	CS/CS	12 shells

SECTION 300
HYDROCRACKING/SEPARATION
CASE 4
TEMPERATURE = 1255°K (1800°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
L-301 A/D	Hydrocracking reactors	4	3.84 m ϕ x 27.44 m (12 ft 7 in. ϕ x 90 ft)	(Metal) 533 (500) 811 (1000)	13.78 (2000)	SA-533 Internal insulation: 20 cm (8 in.)	
				S/T	S/T	S/T	
T-301 A/B	Reactor coolers	2	Total = 1208 m ² (13,000 ft ²)	394/811 (250/1000)	1.72/13.78 (250/2000)	CS/2-1/4 Cr - 1 Mo	
M-350	High-pressure flash drum	1	3.35 m ϕ x 12.20 m (11 ft 0 in. ϕ x 40 ft)	727 (850)	3.45 (500)	SA-533	
M-353 A/B	Atmospheric flash drum and stripper	2	4.09 m ϕ x 10.98 m (13 ft 5 in. ϕ x 36 ft)	727 (850)	0.345 (50)	SA-533	Includes 4 valve trays each
M-355 A/B	Vacuum flash drum	2	7.19 m ϕ x 28.96 m (23 ft 7 in. ϕ x 95 ft)	727 (850)	0.1/vac (15/vac)	SA-533	
P-352 A/B	Atmospheric flash OVHD product pump	2	Total = 0.035 m ³ /sec (557 gpm)	727 (850)	13.78 (2000)	Centrifugal	Motor drive total = 922 kW (1236 hp)
R-350	Flashed vapor booster compressors	1	2.772 kg/sec (22,000 lb/hr) 1843 kW (2472 bhp) MW = 8.7			2-stage centrifugal	Suction = 2.75 MN/m ² , 329°K (400 psig, 132°F) Discharge = 11.37 MN/m ² (1650 psig)
T-352	Atmospheric flash OVHD vapor/product exchanger	1	Total = 799 m ² (8600 ft ²)	705/616 (810/650)	0.345/14.47 (50/2100)	CS/CS	3 shells
T-353	Atmospheric flash CW condenser	1	Total = 1162 m ² (12,500 ft ²)	508/366 (455/200)	0.345/0.517 (50/75)	CS/CS	3 shells
T-354	Vacuum flash feed heater	1	Total = 1627 m ² (17,500 ft ²)	839/727 (1050/850)	2.067/0.345 (300/50)	CS/CS	4 shells
T-355	Vacuum flash OVHD vapor/product exchanger	1	Total = 3792 m ² (40,800 ft ²)	727/616 (850/650)	0.1/vac/0.86 (15/vac/125)	CS/CS	8 shells
T-356	Vacuum flash CW condenser	1	Total = 2789 m ² (30,000 ft ²)	505/366 (450/200)	0.1/vac/0.52 (15/vac/75)	CS/CS	12 shells

SECTION 300
HYDROCRACKING/SEPARATION
CASE 5
TEMPERATURE = 1366°K (2000°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
L-301 A/D	Hydrocracking reactors	4	3.94 m ϕ x 27.44 m (12 ft 11 in. ϕ x 90 ft)	(Metal) 533 (500) 811 (1000)	13.78 (2000)	SA-533 Internal insulation: 20 cm (8 in.)	
				S/T	S/T	S/T	
T-301 A/B	Reactor coolers	2	Total = 1255 m ² (13,500 ft ²)	394/811 (250/1000)	1.72/13.78 (250/2000)	CS/2-1/4 Cr - 1 Mo	
M-350	High-pressure flash drum	1	3.43 m ϕ x 12.20 m (11 ft 3 in. ϕ x 40 ft)	727 (850)	3.45 (500)	SA-533	
M-353 A/B	Atmospheric flash drum and stripper	2	4.17 m ϕ x 10.98 m (13 ft 8 in. ϕ x 36 ft)	727 (850)	0.345 (50)	SA-533	Includes 4 valve trays each
M-355 A/B	Vacuum flash drum	2	7.34 m ϕ x 28.96 m (24 ft 1 in. ϕ x 95 ft)	727 (850)	0.1/vac (15/vac)	SA-533	
P-352 A/B	Atmospheric flash OVHD product pump	2	Total = 0.037 m ³ /sec (581 gpm)	727 (850)	13.78 (2000)	Centrifugal	Motor drive total = 962 kW (1290 hp)
R-350	Flashed vapor booster compressors	1	2.893 kg/sec (22,960 lb/hr) 1924 kW (2580 bhp) MW = 8.7			2-stage centrifugal	Suction = 2.75 MN/m ² , 329°K (400 psig, 132°F) Discharge = 11.37 MN/m ² (1650 psig)
T-352	Atmospheric flash OVHD vapor/product exchanger	1	Total = 837 m ² (9000 ft ²)	705/616 (810/650)	0.345/14.47 (50/2100)	CS/CS	3 shells
T-353	Atmospheric Flash CW condenser	1	Total = 1199 m ² (12,900 ft ²)	508/366 (455/200)	0.345/0.517 (50/75)	CS/CS	3 shells
T-354	Vacuum flash feed heater	1	Total = 1673 m ² (18,000 ft ²)	839/727 (1050/850)	2.067/0.345 (300/50)	CS/CS	4 shells
T-355	Vacuum flash OVHD vapor/product exchanger	1	Total = 3950 m ² (42,500 ft ²)	727/616 (850/650)	0.1/vac/0.86 (15/vac/125)	CS/CS	8 shells
T-356	Vacuum flash CW condenser	1	Total = 2881 m ² (31,000 ft ²)	505/366 (450/200)	0.1/vac/0.52 (15/vac/75)	CS/CS	12 shells

SECTION 400
FLUID COKING
CASE 1
TEMPERATURE = 922°K (1200°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
A-401	Fluid coker	1	2.31 m, 2.8 m, 6.25 m, 2.08 m ϕ , x 36.6 m (7 ft 7 in., 9 ft 2 in., 20 ft 6 in., 6 ft 10 in. ϕ , x 120 ft)	825 (1025)	0.17/0.48 (25/70)	SA-299	
A-402	Fluid coker burner	1	4.75 m ϕ x 6.10 m (15 ft 7 in. ϕ x 20 ft) plus 3.05 m (10 ft) cone	922 (1200)	0.17/0.28 (25/40)	SA-299	With internal monolithic lining
A-403	Coke elutriators	1	1.63 m, 2.90 m ϕ , x 8.23 m (5 ft 4 in., 9 ft 6 in. ϕ , x 27 ft)	922 (1200)	0.24 (35)	SA-299	With internal monolithic lining
M-403	Fluid coker net OVHD product drum	1	1.85 m ϕ x 6.10 m (6 ft 1 in. ϕ x 20 ft)	450 (350)	0.17 (25)	SA-299	
M-404	Air compressor suction drum	1	1.63 m ϕ x 4.57 m (5 ft 4 in. ϕ x 15 ft)	394 (250)	0.103 (15)	SA-299	
P-401 A/B	Fluid coker circulator reflux and spare	2	Total = 0.0202 m ³ /sec (320 gpm)	616 (650)	$\Delta P = 0.41$ ($\Delta P = 60$)	Centrifugal	Motor drive = 16.4 kW (22 hp)
R-401 A/B	Air compressor	2	13.9 kg/sec at 289°K (110,000 lb/hr at 60°F)	311 (100)	$\Delta P = 0.14$ ($\Delta P = 20$)	Centrifugal	Total = 1940 kW (2600 bhp)

SECTION 400
FLUID COKING
CASE 2
TEMPERATURE = 1033°K (1400°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
A-401	Fluid coker	1	3.20 m, 3.84 m, 8.61 m, 2.87 m ϕ , x 36.6 m (10 ft 6 in., 12 ft 7 in., 28 ft 3 in., 9 ft 5 in. ϕ , x 120 ft)	825 (1025)	0.17/0.48 (25/70)	SA-299	
A-402	Fluid coker burner	1	6.53 m ϕ x 6.1 m (21 ft 5 in. ϕ x 20 ft) plus 3.05 m (10 ft) cone	922 (1200)	0.17/0.28 (25/40)	SA-299	With internal monolithic lining
A-403	Coke elutriators	1	2.23 m, 3.96 m ϕ , x 8.23 m (7 ft 4 in., 13 ft ϕ , x 27 ft)	922 (1200)	0.24 (35)	SA-299	With internal monolithic lining
M-403	Fluid coker net OVHD product drum	1	2.54 m ϕ x 6.1 m (8 ft 4 in. ϕ x 20 ft)	450 (350)	0.17 (25)	SA-299	
M-404	Air compressor suction drum	1	2.23 m ϕ x 4.57 m (7 ft 4 in. ϕ x 15 ft)	394 (250)	0.103 (15)	SA-299	
P-401 A/B	Fluid coker circulator reflux and spare	2	Total = 0.038 m ³ /sec (600 gpm)	616 (650)	ΔP = 0.41 (ΔP = 60)	Centrifugal	Motor drive = 31 kW (41 hp)
R-401 A/B	Air compressor	2	26.2 kg/sec at 289°K (208,000 lb/hr at 60°F)	311 (100)	ΔP = 0.14 (ΔP = 20)	Centrifugal	Total = 3672 kW (4925 bhp)

SECTION 400
FLUID COKING
CASE 3
TEMPERATURE = 1144°K (1600°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
A-401	Fluid coker	1	3.28 m, 3.91 m, 8.82 m, 2.45 m ϕ , x 36.6 m (10 ft 9 in., 12 ft 10 in., 28 ft 11 in., 9 ft 8 in. ϕ , x 120 ft)	825 (1025)	0.17/0.48 (25/70)	SA-299	
A-402	Fluid coker burner	1	6.71 m ϕ x 6.1 m (22 ft ϕ x 20 ft) plus 3.05 m (10 ft) cone	922 (1200)	0.17/0.28 (25/40)	SA-299	With internal monolithic lining
A-403	Coke elutriators	1	2.29 m, 4.1 m ϕ , x 8.23 m (7 ft 6 in., 13 ft 5 in. ϕ , x 27 ft)	922 (1200)	0.24 (35)	SA-299	With internal monolithic lining
M-403	Fluid coker net OVHD product drum	1	2.62 m ϕ x 6.1 m (8 ft 7 in. ϕ x 20 ft)	450 (350)	0.17 (25)	SA-299	
M-404	Air compressor suction drum	1	2.24 m ϕ x 4.57 m (7 ft 6 in. ϕ x 15 ft)	394 (250)	0.103 (15)	SA-299	
P-401 A/B	Fluid coker circulator reflux and spare	2	Total = 0.04 m ³ /sec (630 gpm)	616 (650)	$\Delta P = 0.41$ ($\Delta P = 60$)	Centrifugal	Motor drive = 32 kW (43 hp)
R-401 A/B	Air compressor	2	27.47 kg/sec at 289°K (218,000 lb/hr at 60°F)	311 (100)	$\Delta P = 0.14$ ($\Delta P = 20$)	Centrifugal	Total = 3848 kW (5160 bhp)

SECTION 400
FLUID COKING
CASE 4
TEMPERATURE = 1255°K (1800°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
A-401 A/B	Fluid coker	2	2.41 m, 2.87 m, 6.48 m, 2.16 m ϕ , x 36.6 m (7 ft 11 in., 9 ft 5 in., 21 ft 3 in., 7 ft 1 in. ϕ , x 120 ft)	825 (1025)	0.17/0.48 (25/70)	SA-299	
A-402 A/B	Fluid coker burner	2	4.93 m ϕ x 6.1 m (16 ft 2 in. ϕ x 20 ft) plus 3.05 m (10 ft) cone	922 (1200)	0.17/0.28 (25/40)	SA-299	With internal monolithic lining
A-403 A/B	Coke elutriators	2	1.68 m, 3.0 m ϕ , x 8.23 m (5 ft 6 in., 9 ft 10 in. ϕ , x 27 ft)	922 (1200)	0.24 (35)	SA-299	With internal monolithic lining
M-403 A/B	Fluid coker net OVHD product drum	2	1.93 m ϕ x 6.1 m (6 ft 4 in. ϕ x 20 ft)	450 (350)	0.17 (25)	SA-299	
M-404 A/B	Air compressor suction drum	2	1.68 m ϕ x 4.57 m (5 ft 6 in. ϕ x 15 ft)	394 (250)	0.103 (15)	SA-299	
P-401 A/D	Fluid coker circulator reflux and spare	4	Total = 0.043 m ³ /sec (680 gpm)	616 (650)	$\Delta P = 0.41$ ($\Delta P = 60$)	Centrifugal	Motor drive = 35 kW (47 hp)
R-401 A/B	Air compressor	2	29.61 kg/sec at 289°K (235,000 lb/hr at 60°F)	311 (100)	$\Delta P = 0.14$ ($\Delta P = 20$)	Centrifugal	Total = 4146 kW (5560 bhp)

SECTION 400
FLUID COKING
CASE 5
TEMPERATURE = 1366°K (2000°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
A-401 A/B	Fluid coker	2	2.46 m, 2.95 m, 6.61 m, 2.21 m ϕ , x 36.6 m (8 ft 1 in., 9 ft 8 in., 21 ft 8 in., 7 ft 3 in. ϕ , x 120 ft)	825 (1025)	0.17/0.48 (25/70)	SA-299	
A-402 A/B	Fluid coker burner	2	5.03 m ϕ x 6.1 m (16 ft 6 in. ϕ x 20 ft) plus 3.05 m (10 ft) cone	922 (1200)	0.17/0.28 (25/40)	SA-299	With internal monolithic lining
A-403 A/B	Coke elutriators	2	1.73 m, 3.07 m ϕ , x 8.23 m (5 ft 8 in., 10 ft 1 in. ϕ , x 27 ft)	922 (1200)	0.24 (35)	SA-299	With internal monolithic lining
M-403 A/B	Fluid coker net OVHD product drum	2	1.96 m ϕ x 6.1 m (6 ft 5 in. ϕ x 20 ft)	450 (350)	0.17 (25)	SA-299	
M-404 A/B	Air compressor suction drum	2	1.73 m ϕ x 4.57 m (5 ft 8 in. ϕ x 15 ft)	394 (250)	0.103 (15)	SA-299	
P-401 A/D	Fluid coker circulator reflux and spare	4	Total = 0.045 m ³ /sec (710 gpm)	616 (650)	$\Delta P = 0.41$ ($\Delta P = 60$)	Centrifugal	Motor drive = 37 kW (49 hp)
R-401 A/B	Air compressor	2	30.88 kg/sec at 289°K (245,000 lb/hr at 60°F)	311 (100)	$\Delta P = 0.14$ ($\Delta P = 20$)	Centrifugal	Total = 4330 kW (5810 bhp)

SECTION 500
HYDROGASIFICATION/PRODUCT SEPARATION
CASE 1
TEMPERATURE = 922°K (1200°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure (MN/m ² (psig))	Material of Construction or Pump Type	Notes
L-501 A/C	Hydrogasification reactors	3	2.49 m ϕ x 30.49 m (8 ft 2 in. ϕ x 100 ft)	533 (500)	11.37 (1650)	SA-299 Refractory: 17.8 cm (7 in.)	Operating temperature within refractory = 1033°K (1400°F)
M-501	Hydrogasifier feed primary separator	1	1.98 m ϕ x 12.20 m (6 ft 6 in. ϕ x 40 ft)	616 (650)	12.75 (1850)	SA-533	
				S/T	S/T		
T-501	Hydrocracking vapor primary condenser	1	Total = 1069 m ² (11,500 ft ²)	658/714 (725/825)	12.75/12.75 (1850/1850)	C - 1/2 No	3 shells
T-506 A	Hydrogasification effluent/distribution exchanger	1	Total = 1069 m ² (11,500 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	Incoloy 800	3 shells
T-506 B	Hydrogasification effluent/distribution exchanger	1	Total = 1069 m ² (11,500 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	316 SS	3 shells
T-507 A	Hydrogasification effluent/H ₂ exchanger	1	Total = 930 m ² (10,000 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-507 B	Hydrogasification effluent/H ₂ exchanger	1	Total = 930 m ² (10,000 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-508	Hydrogen pre-heater	1	Total = 930 m ² (10,000 ft ²)	589/727 (600/850)	15.16/11.37 (2200/1650)	C - 1/2 No	3 shells
T-509 A	H ₂ /hydrogasification exchanger	1	Total = 660 m ² (7100 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	2 shells
T-509 B	H ₂ /hydrogasification exchanger	1	Total = 660 m ² (7100 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	2 shells
T-510 A	Solvent/hydrogasification exchanger	1	Total = 558 m ² (6000 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	2 shells
T-510 B	Solvent/hydrogasification exchanger	1	Total = 558 m ² (6000 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	2 shells
T-511	Reformer preheater	1	Total = 3579 m ² (38,500 ft ²)	616/672 (650/750)	3.45/10.34 (500/1500)	C - 1/2 No	9 shells
T-512	Hydrogasification effluent cooler	1	Total = 1743 m ² (18,750 ft ²)	366/477 (200/400)	0.69/10.34 (100/1500)	CS/CS	4 shells
A-557 A/B(a)	Aromatics absorbers	2	2.11 m ϕ x 25.30 m (6 ft 11 in. ϕ x 83 ft)	366 (200)	10.51 (1525)	SA-299	20 valve trays (2 pass)
A-558 A/B(a)	Rich oil strippers	2	1.40 m ϕ , 2.06 m ϕ , 17.38 m (4 ft 7 in. ϕ , 6 ft 9 in. ϕ , x 57 ft)	477/630 (400/675)	0.69 (100)	SA-299	20 valve trays
A-559(a)	Stabilizer	1	1.17 m ϕ x 15.24 m (3 ft 10 in. ϕ x 50 ft)	477 (400)	0.86 (125)	SA-299	20 valve trays (2 pass)
M-557(a)	Rich oil flash drum	1	1.47 m ϕ x 7.32 m (4 ft 10 in. ϕ x 24 ft)	366 (200)	1.72 (250)	SA-299	SS wire mesh pad 10 cm (4 in.) thick 0.66 m ϕ (2 ft 2 in. :)
T-569(a)	Lean oil cooler	1	Total = 702 m ² (7550 ft ²)	339/366 (150/200)	11.03/0.52 (1600/75)	CS/CS	2 shells
T-571(a)	Absorber intercoolers	1	Total = 1701 m ² (18,300 ft ²)	366/366 (200/200)	10.51/0.52 (1525/75)	CS/CS	4 shells
T-572(a)	Lean oil exchanger	1	Total = 558 m ² (6000 ft ²)	616/505 (650/450)	0.69/1.03 (100/150)	CS/CS	2 shells
P-555 A/B(a)	Lean oil pumps	2	Total = 0.11 m ³ /sec (1734 gpm)	422 (300)	$\Delta P = 10.34$ ($\Delta P = 1500$)	Centrifugal	Motor total power = 1595 kW (2139 hp)

(a) Equipment included in gas cleanup block shown in hydrogasification unit.

REPRODUCIBILITY OF THE
ORIGINAL PAGE IS POOR

SECTION 500
HYDROGASIFICATION/PRODUCT SEPARATION
CASE 2
TEMPERATURE = 1033°K (1400°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
L-501	Hydrogasification reactors	3	3.40 m ϕ x 30.49 m (11 ft 2 in. ϕ x 100 ft)	533 (500)	11.37 (1650)	SA-299 Refractory: 17.8 cm (7 in.) SA-533	Operating temperature within refractory = 1033°K (1400°F)
M-501	Hydrogasifier feed primary condenser	1	2.72 m ϕ x 12.20 m (8 ft 11 in. ϕ x 40 ft)	616 (650)	12.75 (1850)		
				S/T	S/T		
T-501	Hydrocracking vapor primary separator	1	Total = 2036 (21,900 ft ²)	658/714 (725/825)	12.75/12.75 (1850/1850)	C - 1/2 Mo	5 shells
T-506 A	Hydrogasification effluent/distribution exchanger	1	Total = 2036 m ² (21,900 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	Incoloy 800	5 shells
T-506 B	Hydrogasification effluent/distribution exchanger	1	Total = 2036 m ² (21,900 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	316 SS	5 shells
T-507 A	Hydrogasification effluent/H ₂ exchanger	1	Total = 1780 m ² (19,150 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-507 B	Hydrogasification effluent/H ₂ exchanger	1	Total = 1780 m ² (19,150 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	4 shells
T-508	Hydrogen preheater	1	Total = 1780 m ² (19,150 ft ²)	589/727 (600/850)	15.16/11.37 (2200/1650)	C - 1/2 Mo	5 shells
T-509 A	H ₂ /hydrogasification exchanger	1	Total = 1259 m ² (13,550 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-509 B	H ₂ /hydrogasification exchanger	1	Total = 1259 m ² (13,550 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-510 A	Solvent/hydrogasification exchanger	1	Total = 1032 m ² (11,100 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-510 B	Solvent/hydrogasification exchanger	1	Total = 1032 m ² (11,100 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	2 shells
T-511	Reformer preheater	1	Total = 6776 m ² (72,900 ft ²)	616/672 (650/750)	3.45/10.34 (500/1500)	C - 1/2 Mo	9 shells
T-512	Hydrogasification effluent cooler	1	Total = 3304 m ² (33,550 ft ²)	366/477 (200/400)	0.69/10.34 (100/1500)	CS/CS	6 shells
A-557 A/B(a)	Aromatics absorbers	2	2.90 m ϕ x 25.30 m (9 ft 6 in. ϕ x 83 ft)	366 (200)	10.51 (1525)	SA-299	20 valve trays (2 pass)
A-558 A/B(a)	Rich oil strippers	2	1.93 m ϕ , 2.82 m ϕ , x 17.38 m (6 ft 4 in. ϕ , 9 ft 3 in. ϕ , x 57 ft)	477/630 (400/675)	0.69 (100)	SA-299	20 valve trays
A-559(a)	Stabilizer	1	1.60 m ϕ x 15.24 m (5 ft 3 in. ϕ x 50 ft)	477 (400)	0.86 (125)	SA-299	20 valve trays (2 pass)
M-557(a)	Rich oil flash drum	1	2.02 m ϕ x 7.32 m (6 ft 8 in. ϕ x 24 ft)	366 (200)	1.72 (250)	SA-299	SS wire mesh pad 10 cm (4 in.) thick 0.90 m (3 ft ϕ)
T-569(a)	Lean oil cooler	1	Total = 1329 m ² (14,300 ft ²)	339/366 (150/200)	11.03/0.52 (1600/75)	CS/CS	2 shells
T-571(a)	Absorber intercoolers	1	Total = 3321 m ² (34,650 ft ²)	366/366 (200/200)	10.51/0.52 (1525/75)	CS/CS	4 shells
T-572(a)	Lean oil exchanger	1	Total = 1022 m ² (11,000 ft ²)	616/505 (650/450)	0.69/1.03 (100/150)	CS/CS	2 shells
P-555 A/C(a)	Lean oil pumps	3	Total = 0.2070 m ³ /sec (3282 ppm)	422 (300)	NP = 10.34 (NP = 1500)	Centrifugal	Motor total power = 3920 kW (4050 hp)

(a) Equipment included in gas cleanup block shown in hydrogasification unit.

SECTION 500
HYDROGASIFICATION/PRODUCT SEPARATION
CASE 3
TEMPERATURE = 1144°K (1600°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure (MN/m ² (psig))	Material of Construction or Pump Type	Notes
L-501	Hydrogasification reactors	3	3.51 m ϕ x 30.48 m (11 ft 6 in. ϕ x 100 ft)	533 (500)	11.37 (1650)	SA-299 Refractory: 17.8 cm (7 in.)	Operating temperature within refractory = 1033°K (1400°F)
M-501	Hydrogasifier feed primary separator	1	2.77 m ϕ x 12.20 m (9 ft 1 in. ϕ x 40 ft)	616 (650)	12.75 (1850)	SA-533	
				S/T	S/T		
T-501	Hydrocracking vapor primary condenser	1	Total = 2140 m ² (23,000 ft ²)	658/714 (725/825)	12.75/12.75 (1850/1850)	C - 1/2 Ho	5 shells
T-506 A	Hydrogasification effluent/distribution exchanger	1	Total = 2140 m ² (23,000 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	Incoloy 800	5 shells
T-506 B	Hydrogasification effluent/distribution exchanger	1	Total = 2140 m ² (23,000 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	316 SS	5 shells
T-507 A	Hydrogasification effluent/H ₂ exchanger	1	Total = 1860 m ² (20,000 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	7 shells
T-507 B	Hydrogasification effluent/H ₂ exchanger	1	Total = 1860 m ² (20,000 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	7 shells
T-508	Hydrogen preheater	1	Total = 1860 m ² (20,000 ft ²)	589/727 (600/850)	15.16/11.37 (2200/1650)	C - 1/2 Ho	7 shells
T-509 A	H ₂ /hydrogasification exchanger	1	Total = 1320 m ² (14,200 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-509 B	H ₂ /hydrogasification exchanger	1	Total = 1320 m ² (14,200 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-510 A	Solvent/hydrogasification exchanger	1	Total = 1080 m ² (11,600 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	4 shells
T-510 B	Solvent/hydrogasification exchanger	1	Total = 1080 m ² (11,600 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	4 shells
T-511	Reformer preheater	1	Total = 7100 m ² (76,400 ft ²)	616/672 (650/750)	3.45/10.34 (500/1500)	C - 1/2 Ho	17 shells
T-512	Hydrogasification effluent cooler	1	Total = 3460 m ² (37,250 ft ²)	366/477 (200/400)	0.69/10.34 (100/1500)	CS/CS	8 shells
A-557 A/B(a)	Aromatics absorbers	2	2.97 m ϕ x 25.3 m (9 ft 9 in. ϕ x 83 ft)	366 (200)	10.51 (1525)	SA-299	20 valve trays (2 pass)
A-558 A/B(a)	Rich oil strippers	2	1.98 m ϕ , 2.90 m ϕ , x 17.38 m (6 ft 6 in. ϕ , 9 ft 6 in. ϕ , x 57 ft)	477/630 (400/675)	0.69 (100)	SA-299	20 valve trays
A-559(a)	Stabilizer	1	1.63 m ϕ x 15.24 m (5 ft 4 in. ϕ x 50 ft)	477 (400)	0.86 (125)	SA-299	20 valve trays (2 pass)
M-557(a)	Rich oil flash drum	1	2.08 m ϕ x 7.32 m (6 ft 10 in. ϕ x 24 ft)	366 (200)	1.72 (250)	SA-299	SS wire mesh pad 10 cm (4 in.) thick
T-569(a)	Lean oil cooler	1	Total = 1395 m ² (15,000 ft ²)	339/366 (150/200)	11.03/0.52 (1600/75)	CS/CS	5 shells
T-571(a)	Absorber intercoolers	1	Total = 3375 m ² (36,300 ft ²)	366/366 (200/200)	10.51/0.52 (1525/75)	CS/CS	8 shells
T-572(a)	Lean oil exchanger	1	Total = 1070 m ² (11,500 ft ²)	616/505 (650/450)	0.69/1.03 (100/150)	CS/CS	4 shells
P-555 A/D(a)	Lean oil pumps	4	Total = 0.221 m ³ /sec (3500 gpm)	422 (300)	AP = 10.34 (AP = 1500)	Centrifugal	Motor total power = 3210 kW (4300 hp)

(a) Equipment included in gas cleanup block shown in hydrogasification unit.

SECTION 500
HYDROGASIFICATION/PRODUCT SEPARATION
CASE 4
TEMPERATURE = 1255°K (1800°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [N/m ² (psig)]	Material of Construction or Pump Type	Notes
L-501	Hydrogasification reactors	3	3.63 m ϕ x 30.48 m (11 ft 11 in. ϕ x 100 ft)	533 (500)	11.37 (1650)	SA-299 Refractory: 17.8 cm (7 in.)	Operating temperature within refractory = 1033°K (1400°F)
M-501	Hydrogasifier feed primary separator	1	2.90 m ϕ x 12.20 m (9 ft 6 in. ϕ x 40 ft)	616 (650)	12.75 (1850)	SA-533	
				S/T	S/T		
T-501	Hydrogasifier vapor primary condenser	1	Total = 2325 m ² (25,000 ft ²)	658/714 (725/825)	12.75/12.75 (1850/1850)	C - 1/2 Mo	6 shells
T-506 A	Hydrogasification effluent/distribution exchanger	1	Total = 2325 m ² (25,000 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	Incoloy 800	6 shells
T-506 B	Hydrogasification effluent/distribution exchanger	1	Total = 2325 m ² (25,000 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	316 SS	6 shells
T-507 A	Hydrogasification effluent/H ₂ exchanger	1	Total = 2000 m ² (21,500 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-507 B	Hydrogasification effluent/H ₂ exchanger	1	Total = 2000 m ² (21,500 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-508	Hydrogen preheater	1	Total = 2000 m ² (21,500 ft ²)	589/727 (609/850)	15.16/11.37 (2200/1650)	C - 1/2 Mo	5 shells
T-509 A	H ₂ /hydrogasification exchanger	1	Total = 1420 m ² (15,300 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-509 B	H ₂ /hydrogasification exchanger	1	Total = 1420 m ² (15,300 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-510 A	Solvent/hydrogasification exchanger	1	Total = 1160 m ² (12,500 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-510 B	Solvent/hydrogasification exchanger	1	Total = 1160 m ² (12,500 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-511	Reformer preheater	1	Total = 7650 m ² (82,300 ft ²)	616/672 (650/750)	1.45/10.34 (500/1500)	C - 1/2 Mo	18 shells
T-512	Hydrogasification effluent cooler	1	Total = 3720 m ² (40,000 ft ²)	366/477 (200/400)	0.69/10.34 (100/1500)	CS/CS	9 shells
A-557 A/B(a)	Aromatics absorbers	2	3.10 m ϕ x 25.3 m (10 ft 2 in. ϕ x 83 ft)	366 (200)	10.51 (1525)	SA-299	20 valve trays (2 pass)
A-558 A/B(a)	Rich oil strippers	2	2.03 m ϕ , 3.0 m ϕ , x 17.38 m (6 ft 8 in. ϕ , 9 ft 10 in. ϕ , x 57 ft)	477/630 (400/675)	0.69 (100)	SA-299	20 valve trays
A-559(a)	Stabilizer	1	1.68 m ϕ x 15.24 m (5 ft 6 in. ϕ x 50 ft)	477 (400)	0.86 (125)	SA-299	20 valve trays (2 pass)
M-557(a)	Rich oil flash drum	1	2.16 m ϕ x 7.32 m (7 ft 1 in. ϕ x 24 ft)	366 (200)	1.72 (250)	SA-299	SS wire mesh pad 10 cm (4 in.) thick
T-569(a)	Lean oil cooler	1	Total = 1490 m ² (16,000 ft ²)	339/366 (150/200)	11.03/0.52 (1600/75)	CS/CS	4 shells
T-571(a)	Absorber intercoolers	1	Total = 3625 m ² (39,000 ft ²)	366/366 (200/200)	10.51/0.52 (1525/75)	CS/CS	9 shells
T-572(a)	Lean oil exchanger	1	Total = 1160 m ² (12,500 ft ²)	616/505 (650/450)	0.69/1.03 (100/1500)	CS/CS	3 shells
P-555 A/D(a)	Lean oil pumps	4	Total = 0.234 m ³ /sec (3708 gpm)	422 (300)	P = 10.34 (P = 1500)	Centrifugal	Motor total power = 3410 kW (4574 hp)

(a) Equipment included in gas cleanup block shown in hydrogasification unit.

SECTION 500
HYDROGASIFICATION/PRODUCT SEPARATION
CASE 5
TEMPERATURE = 1366°K (2000°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
L-501	Hydrogasification reactors	3	3.71 m ϕ x 30.48 m (12 ft 2 in. ϕ x 100 ft)	533 (500)	11.37 (1650)	SA-299 Refractory: 17.8 cm (7 in.)	Operating temperature within refractory = 1033°K (1400°F)
M-501	Hydrogasifier feed primary separator	1	2.95 m ϕ x 12.20 m (9 ft 8 in. ϕ x 40 ft)	616 (650)	12.75 (1850)	SA-533	
				S/T	S/T		
T-501	Hydrocracking vapor primary condenser	1	Total = 2420 m ² (26,000 ft ²)	658/714 (725/825)	12.75/12.75 (1850/1850)	C - 1/2 Mo	6 shells
T-506 A	Hydrogasification effluent/distribution exchanger	1	Total = 2420 m ² (26,000 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	Incoloy 800	6 shells
T-506 B	Hydrogasification effluent/distribution exchanger	1	Total = 2420 m ² (26,000 ft ²)	950/1061 (1250/1450)	11.71/11.37 (1700/1650)	316 SS	6 shells
T-507 A	Hydrogasification effluent/H ₂ exchanger	1	Total = 2090 m ² (22,500 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-507 B	Hydrogasification effluent/H ₂ exchanger	1	Total = 2090 m ² (22,500 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	5 shells
T-508	Hydrogen preheater	1	Total = 2090 m ² (22,500 ft ²)	589/727 (600/850)	15.16/11.37 (2200/1650)	C - 1/2 Mo	5 shells
T-509 A	H ₂ /hydrogasification exchanger	1	Total = 1490 m ² (16,000 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	4 shells
T-509 B	H ₂ /hydrogasification exchanger	1	Total = 1490 m ² (16,000 ft ²)	950/1061 (1250/1450)	15.16/11.37 (2200/1650)	Incoloy 800	4 shells
T-510 A	Solvent/hydrogasification exchanger	1	Total = 1210 m ² (13,000 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-510 B	Solvent/hydrogasification exchanger	1	Total = 1210 m ² (13,000 ft ²)	755/1061 (900/1450)	15.16/11.37 (2200/1650)	Incoloy 800	3 shells
T-511	Reformer preheater	1	Total = 7990 m ² (86,000 ft ²)	616/672 (650/750)	3.45/10.34 (500/1500)	C - 1/2 Mo	20 shells
T-512	Hydrogasification effluent cooler	1	Total = 3900 m ² (42,000 ft ²)	366/477 (200/400)	0.69/10.34 (100/1500)	CS/CS	10 shells
A-557 A/B(a)	Aromatics absorbers	2	3.15 m ϕ x 25.3 m (10 ft 4 in. ϕ x 83 ft)	366 (200)	10.51 (1525)	SA-299	20 valve trays (2 pass)
A-558 A/B(a)	Rich oil strippers	2	2.08 m ϕ , 3.05 m ϕ , x 17.38 m (6 ft 10 in. ϕ , 10 ft ϕ , x 57 ft)	477/630 (400/675)	0.69 (100)	SA-299	20 valve trays
A-559(a)	Stabilizer	1	1.73 m ϕ x 15.24 m (5 ft 8 in. ϕ x 50 ft)	477 (400)	0.86 (125)	SA-299	20 valve trays (2 pass)
M-557(a)	Rich oil flash drum	1	2.21 m ϕ x 7.32 m (7 ft 3 in. ϕ x 24 ft)	366 (200)	1.72 (250)	SA-299	SS wire mesh pad 10 cm (4 in.) thick
T-569(a)	Lean oil cooler	1	Total = 1580 m ² (17,000 ft ²)	339/366 (150/200)	11.03/0.52 (1600/75)	CS/CS	4 shells
T-571(a)	Absorber intercoolers	1	Total = 3810 m ² (41,000 ft ²)	366/366 (200/200)	10.51/0.52 (1525/75)	CS/CS	10 shells
T-572(a)	Lean oil exchanger	1	Total = 1210 m ² (13,000 ft ²)	616/505 (650/450)	0.69/1.03 (100/150)	CS/CS	4 shells
P-555 A/B(a)	Lean oil pumps	4	Total = 0.244 m ³ /sec (3870 gpm)	422 (300)	NP = 10.34 (NP = 1500)	Centrifugal	Motor total power = 3560 kW (4773 hp)

(a) Equipment included in gas cleanup block shown in hydrogasification unit.

SECTION 600
HYDROGEN PRODUCTION AND POWER GENERATION
CASE 1
TEMPERATURE = 922°K (1200°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
B-601 A/F	Reformer steam generator super-heater	6	20,450 m ² total (220,000 ft ²)	625/633 (665/680)	3.38/5.96 (490/865)	CS/CS	
P-601 A/H	Power steam BFW pump	8	1.181 m ³ /sec (18,720 gpm)	561 (550)	22.15 (3215)	Centrifugal	Total = 22.0 MW (29,500 hp) ΔP = 16.6 MN/m ² (2405 psi)
T-601 A/F	Reformer feed/effluent exchanger	6	17,850 m ² total (192,000 ft ²)	936/830 (1225/1035)	2.52/3.20 (365/465)	316 SS/316 SS	
T-602 A/F	Power steam generator BFW preheater	6	7160 m ² total (77,000 ft ²)	722/600 (840/620)	2.52/22.15 (365/3215)	316 SS/ Incoloy 800	
T-606	Power steam reheater	0	—	—	—	—	None in case 1
TU-603 A	High-pressure turbine generator	1	66 MW (88,500 hp)	730 (855)	11.96 (1735)		Exhaust = 5.51 MN/m ² (800 psi)
TU-603 B	Low-pressure turbine generator	1	165 MW (221,300 hp)	625 (665)	5.62 (815)		Exhaust = 0.20 MN/m ² (29 psi)

SECTION 600
HYDROGEN PRODUCTION AND POWER GENERATION
CASE 2
TEMPERATURE = 1033°K (1400°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
B-601 A/F	Reformer steam generator super-heater	6	34,400 m ² total (370,000 ft ²)	625/681 (665/765)	4.24/5.27 (615/765)	CS/CS	
P-601 A/F	Power steam BFW pump	6	0.936 m ³ /sec (14,840 gpm)	561 (550)	22.15 (3215)	Centrifugal	17 MW (22,800 hp) ΔP = 17.2 MN/m ² (2500 psi)
T-601 A/F	Reformer feed/effluent exchanger	6	26,490 m ² total (285,000 ft ²)	881/830 (1125/1035)	2.17/4.0 (315/580)	316 SS/316 SS	
T-602 A/F	Power steam generator BFW preheater	6	20,450 m ² total (220,000 ft ²)	694/609 (790/636)	2.31/22.5 (335/3265)	316 SS/Incoloy 800	
T-606	Power steam reheater	6	5600 m ² total (60,000 ft ²)	958/806 (1265/990)	2.31/11.47 (335/1665)	316 SS/Incoloy 800	
TU-603 A	High-pressure turbine generator	0	--	--	--	--	None in case 2
TU-603 B	Low-pressure turbine generator	1	100 MW (134,100 hp)	Steam = 667 (741)	Steam = 4.86 (705)		Exhaust = 0.009 MN/m ² (1.3 psi)

SECTION 600
HYDROGEN PRODUCTION AND POWER GENERATION
CASE 3
TEMPERATURE = 1144°K (1600°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
B-601 A/F	Reformer steam generator super-heater	6	22,800 m ² total (245,000 ft ²)	625/714 (665/825)	4.58/6.13 (665/890)	CS/2-1/4 Cr - 1 Mo	
P-601 A/F	Power steam BFW pump	6	0.897 m ³ /sec (14,220 gpm)	578 (580)	22.15 (3215)	Centrifugal	16 MW (21,400 hp) ΔP = 16.2 MN/m ² (2350 psi)
T-601 A/F	Reformer feed/effluent exchanger	6	28,600 m ² total (308,000 ft ²)	886/853 (1135/1075)	2.41/4.38 (350/635)	316 SS/316 SS	
T-602 A/F	Power steam generator BFW preheater	6	13,300 m ² total (143,000 ft ²)	733/622 (860/660)	2.41/22.5 (350/3265)	316 SS/Incoloy 800	
T-606	Power steam reheater	6	5300 m ² total (57,000 ft ²)	1011/797 (1360/975)	2.41/9.92 (350/1440)	316 SS/Incoloy 800	
TU-603 A	High-pressure turbine generator	1	114 MW (153,000 hp)	783 (950)	9.72 (1410)		
TU-603 B	Low-pressure turbine generator	0	--	--	--	--	None in case 3

SECTION 600
HYDROGEN PRODUCTION AND POWER GENERATION
CASE 4
TEMPERATURE = 1255°K (1800°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
B-601 A/F	Reformer steam generator super-heater	6	19,900 m ² total (214,000 ft ²)	625/652 (665/713)	4.75/6.48 (690/940)	CS/2-1/4 Cr - 1 Mo	
P-601 A/F	Power steam BFW pump	6	0.728 m ³ /sec (11,540 gpm)	561 (550)	22.15 (3215)	Centrifugal	14 MW (18,800 hp) ΔP = 16.02 MN/m ² (2325 psi)
T-601 A/F	Reformer feed/effluent exchanger	6	6990 m ² total (75,200 ft ²)	853/978 (1075/1300)	4.31/3.38 (625/490)	316 SS/ Incoloy 800	
T-602 A/F	Power steam generator BFW preheater	6	2300 m ² total (24,800 ft ²)	783/608 (950/635)	3.38/22.2 (490/3222)	316 SS/ Incoloy 800	
T-606	Power steam reheater	6	2200 m ² total (23,700 ft ²)	1067/853 (1460/1075)	3.38/6.48 (490/940)	316 SS/ Incoloy 800	Shell insulation
TU-603 A	High-pressure turbine generator	1	50 MW (67,050 hp)	Steam = 696 (793)	Steam = 9.48 (1375)		
TU-603 B	Low-pressure turbine generator	1	33 MW (44,250 hp)	Steam = 836 (1045)	Steam = 6.13 (890)		

SECTION 600
HYDROGEN PRODUCTION AND POWER GENERATION
CASE 5
TEMPERATURE = 1366°K (2000°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
B-601 A/F	Reformer steam generator super-heater	6	18,600 m ² total (200,000 ft ²)	625/658 (665/725)	5.27/6.82 (765/990)	CS/CS	
P-601 A/F	Power steam BFW pump	6	0.686 m ³ /sec (10,880 gpm)	575 (575)	22.15 (3215)	Centrifugal	13 MW (17,400 hp) ΔP = 15.7 MN/m ² (2280 psig)
T-601 A/F	Reformer feed/effluent exchanger	6	4140 m ² total (44,500 ft ²)	853/1072 (1075/1470)	5.62/3.96 (815/575)	316 SS/Incoloy 800	
T-602 A/F	Power steam generator BFW preheater	6	5340 m ² total (51,600 ft ²)	881/650 (1125/710)	3.96/22.15 (575/3215)	316 SS/316 SS	
T-606	Power steam reheater	6	1460 m ² total (15,600 ft ²)	1144/853 (1600/1075)	3.96/6.82 (575/990)	316 SS/Incoloy 800	Shell insulation
TU-603 A	High-pressure turbine generator	1	59 MW (79,120 hp)	Steam = 694 (789)	Steam = 9.20 (1335)		
TU-603 B	Low-pressure turbine generator	1	57 MW (76,440 hp)	Steam = 839 (1050)	Steam = 6.44 (935)		

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SECTION 700
HYDROGEN PURIFICATION AND COMPRESSION
CASE 1
TEMPERATURE = 922°K (1200°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
L-701	First-stage shift converter (HTS)	0	--	--	--	--	None in case 1
L-702 A/I	Second-stage shift converter (LTS)	9	4.65 m ϕ x 15.24 m (15 ft 3 in. ϕ x 50 ft)	525 (485)	2.34 (340)	SA-299 Liner: 316 SS	LTS catalyst bed = 12.2 m (40 ft)
L-703 A/F	Methanator	6	3.20 m ϕ x 9.15 m (10 ft 6 in. ϕ x 30 ft)	544 (520)	1.96 (285)	SA-299	Methanation catalyst bed = 6.1 m (20 ft)
A-701 A/F	CO ₂ absorber	6	4.06 m ϕ x 38.11 m (13 ft 4 in. ϕ x 125 ft)	397 (255)	2.17 (315)	SA-299	8.9 cm (3-1/2 in.) Intalox saddle plastic rings
A-702 A/F	CO ₂ stripper	6	5.95 m ϕ x 35.06 m (19 ft 6 in. ϕ x 115 ft)	411 (280)	0.276 (40)	SA-299	5 cm (2 in.) porcelain Raschig rings
A-703	Process gas dryer	0	--	--	--	--	None in case 1
A-704	Solvent dryer	0	--	--	--	--	None in case 1
T-701 A/F	CO ₂ stripper reboiler	6	Total = 8460 m ² (91,000 ft ²)	411/458 (280/365)	0.276/2.17 (40/315)	CS/316 SS	3 shells each
T-702 A/F	BFW heater	6	Total = 14,500 m ² (156,000 ft ²)	547/491 (525/425)	2.34/4.58 (340/665)	316 SS/ 316 SS	4 shells each
T-707 A/F	CO ₂ stripper condenser	6	Total = 7900 m ² (85,000 ft ²)	408/311 (275/100)	0.276/0.79 (40/115)	CS/CS	3 shells each
P-701 A/J	Solvent pump (lean solution or high-pressure solvent)	10	Total = 1795 m ³ /sec (28,450 gpm)	422 (300)	ΔP = 2.24 (325)	Centrifugal	Total = 8874 kW (11,900 hp)
P-702	Low-pressure solvent pump	0	--	--	--	--	None in case 1
R-701	Hydrogen compressor (process gas compressor)	1	115 MW (154,200 hp)	422 (300)	ΔP = 3.79 (550)	Centrifugal	19.6 kg-mole/sec (985,000 scfm)
R-702	Hydrogen booster compressor	1	28 MW (37,600 hp)	367 (200)	ΔP = 8.61 (1250)	Centrifugal	14.29 kg-mole/sec (716,000 scfm)

SECTION 700
HYDROGEN PURIFICATION AND COMPRESSION
CASE 2
TEMPERATURE = 1033°K (1400°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
L-701	First-stage shift converter (HTS)	0	--	--	--	--	None in case 2
L-702 A/I	Second-stage shift converter (LTS)	9	4.88 m ϕ x 15.24 m (16 ft ϕ x 50 ft)	567 (560)	2.34 (340)	SA-299 Liner: 316 SS	LTS catalyst bed = 12.2 m (40 ft)
L-703 A/F	Methanator	6	3.66 m ϕ x 9.14 m (12 ft ϕ x 30 ft)	581 (550)	1.96 (285)	SA-299	Ni catalyst bed = 6.1 m (20 ft)
A-701 A/F	CO ₂ absorber	6	5.18 m ϕ x 38.1 m (17 ft ϕ x 125 ft)	397 (255)	2.17 (315)	SA-299	5 cm (2 in.) porcelain Raschig ring packing
A-702 A/F	CO ₂ stripper	6	7.62 m ϕ x 35.1 m (25 ft ϕ x 115 ft)	411 (280)	0.276 (40)	SA-299	5 cm (2 in.) porcelain Raschig ring packing
A-703	Process gas dryer	0	--	--	--	--	None in case 2
A-704	Solvent dryer	0	--	--	--	--	None in case 2
T-701 A/F	CO ₂ stripper reboiler	6	Total = 25,100 m ² (270,000 ft ²)	411/458 (280/365)	0.276/2.17 (40/315)	CS/316 SS	5 shells each
T-702 A/F	BFW heater	6	Total = 12,100 m ² (130,000 ft ²)	541/492 (525/425)	2.34/4.58 (340/665)	316 SS/ 316 SS	3 shells each
T-707 A/F	CO ₂ stripper condenser	6	Total = 13,000 m ² (140,000 ft ²)	408/311 (275/100)	0.276/0.79 (40/115)	CS/CS	3 shells each
P-701 A/T	Solvent pump (lean solution or high-pressure solvent)	20	Total = 3.79 m ³ /sec (60,000 gpm)	422 (300)	2.52 (365)	Centrifugal	12.8 MW (19,200 hp)
P-702	Low-pressure solvent pump	0	--	--	--	--	None in case 2
R-701	Hydrogen compressor (process gas compressor)	1	210 MW (280,000 hp)	422 (300)	ΔP = 8.79 (1275)	Centrifugal	25.6 kg-mole/sec (1,290,000 scfm)
R-702	Hydrogen booster compressor	1	17 MW (22,000 hp)	367 (200)	ΔP = 3.55 (515)	Centrifugal	13.3 kg-mole/sec (670,000 scfm)

SECTION 700
HYDROGEN PURIFICATION AND COMPRESSION
CASE 3
TEMPERATURE = 1144°K (1600°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
L-701 A/I	First-stage shift converter (HTS)	9	3.35 m ϕ x 9.15 m (11 ft ϕ x 30 ft)	733 (860)	2.41 (350)	SA-533	HTS catalyst bed = 6.1 m (20 ft)
L-702 A/I	Second-stage shift converter (LTS)	9	4.88 m ϕ x 15.2 m (16 ft ϕ x 50 ft)	522 (480)	2.41 (350)	SA-299 Liner: 316 SS	LTS catalyst bed = 12.2 m (40 ft)
L-703 A/F	Methanator	6	3.78 m ϕ x 9.15 m (12 ft 5 in. ϕ x 30 ft)	594 (610)	2.03 (295)	SA-299	Ni catalyst bed = 6.1 m (20 ft)
A-701 A/F	CO ₂ absorber	6	5.46 m ϕ x 38.1 m (17 ft 10 in. ϕ x 125 ft)	397 (255)	2.17 (315)	SA-299	5 cm (2 in.) porcelain Raschig rings
A-702 A/F	CO ₂ stripper	6	8.0 m ϕ x 35.1 m (26 ft 3 in. ϕ x 115 ft)	411 (280)	0.276 (40)	SA-299	5 cm (2 in.) porcelain Raschig rings
A-703	Process gas dryer	0	--	--	--	--	None in case 3
A-704	Solvent dryer	0	--	--	--	--	None in case 3
T-701A A/F	CO ₂ stripper reboiler	6	Total = 5440 m ² (58,600 ft ²)	411/444 (280/340)	0.276/2.34 (40/340)	CS/316 SS	2 shells each
T-701B A/F	CO ₂ stripper reboiler	6	Total = 24,600 m ² (265,000 ft ²)	411/444 (280/340)	0.276/0.62 (40/90)	CS/CS	10 shells each
T-702 A/F	BFW heater	6	Total = 8830 m ² (95,000 ft ²)	536/492 (505/425)	2.34/4.93 (340/715)	316 SS/ 316 SS	4 shells each
T-707 A/F	CO ₂ stripper condenser	6	Total = 14,400 m ² (155,000 ft ²)	408/311 (275/100)	0.276/0.79 (40/115)	CS/CS	6 shells each
P-701 A/T	Solvent pump (lean solution or high-pressure solvent)	20	Total = 3.28 m ³ /sec (51,950 gpm)	422 (300)	2.52 (365)	Centrifugal	12.3 MW (16,470 hp)
P-702	Low-pressure solvent pump	0	--	--	--	--	None in case 3
R-701	Hydrogen compressor (process gas compressor)	1	220 MW (293,000 hp)	422 (300)	ΔP = 8.68 (1260)	Centrifugal	27.5 kg-mole/sec (1,380,000 scfm)
R-702	Hydrogen booster compressor	1	18 MW (23,200 hp)	366 (200)	ΔP = 3.55 (515)	Centrifugal	14.4 kg-mole/sec (720,000 scfm)

SECTION 700
HYDROGEN PURIFICATION AND COMPRESSION
CASE 4
TEMPERATURE = 1255°K (1800°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
L-701 A/I	First-stage shift converter (HTS)	9	3.90 m ϕ x 16.8 m (12 ft 10 in. ϕ x 55 ft)	678 (760)	3.14 (455)	SA-533 Liner: 316 SS	HTS catalyst bed = 13.7 m (45 ft)
L-702 A/I	Second-stage shift converter (LTS)	9	5.46 m ϕ x 19.2 m (17 ft 11 in. ϕ x 63 ft)	547 (525)	3.14 (455)	SA-299 Liner: 316 SS	LTS catalyst bed = 16.2 m (53 ft)
L-703 A/F	Methanator	6	1.52 m ϕ x 9.14 m (5 ft 0 in. ϕ x 30 ft)	611 (640)	11.13 (1615)	SA-299	Ni catalyst bed = 6.1 m (20 ft)
A-701 A/F	CO ₂ absorber	6	4.39 m ϕ x 27.4 m (14 ft 5 in. ϕ x 90 ft)	331 (135)	10.9 (1580)	SA-299	41 sieve trays
A-702 A/F	CO ₂ stripper	6	4.39 m ϕ x 22.9 m (14 ft 5 in. ϕ x 75 ft)	331 (135)	0.276 (40)	SA-299	27 sieve trays
A-703	Process gas dryer	6	4.11 m ϕ x 6.1 m (13 ft 6 in. ϕ x 20 ft)	331 (135)	4.0 (580)	SA-299	5 sieve trays
A-704	Solvent dryer	6	Top: 4.39 m ϕ x 6.1 m (14 ft 5 in. ϕ x 20 ft) Bottom: 2.05 m ϕ x 3.66 m (6 ft 9 in. ϕ x 12 ft)	478 (400)	0.276 (40)	SA-299	Top: 5 bubble cup trays Bottom: 2 bubble cup trays
T-701	CO ₂ stripper reboiler	0	—	—	—	—	None in case 4
T-702A A/F	BFW heater	6	Total = 6350 m ² (68,400 ft ²)	547/492 (525/425)	3.34/5.27 (485/765)	2-1/4 in. - 1 Mo/ 2-1/4 in. - 1 Mo	2 shells each
T-702B A/F	BFW heater	6	Total = 2850 m ² (30,700 ft ²)	458/436 (365/325)	3.34/5.27 (485/765)	316 SS/ 316 SS	
T-707	CO ₂ stripper condenser	0				—	None in case 4
P-701	Solvent pump (lean solution or high-pressure solvent)	27	Total = 5.1 m ³ /sec (80,800 gpm)	331 (135)	ΔP = 5.13 (745)	Centrifugal	32 MW (42,900 hp)
P-702	Low-pressure solvent pump	27	Total = 5.1 m ³ /sec (80,800 gpm)	331 (135)	ΔP = 5.41 (785)	Centrifugal	33 MW (44,300 hp)
R-701	Hydrogen compressor (process gas compressor)	1	327 MW (364,400 hp)	400 (260)	ΔP = 7.75 (1125)	Centrifugal	38.9 kg-mole/sec (1,950,000 scfm)
R-702	Hydrogen booster compressor	1	19 MW (25,500 hp)	372 (210)	ΔP = 3.55 (515)	Centrifugal	14.1 kg-mole/sec (705,000 scfm)

SECTION 700
HYDROGEN PURIFICATION AND COMPRESSION
CASE 5
TEMPERATURE = 1366°K (2000°F)

Item No.	Description	Quantity	Size, Capacity	Design Temperature [°K (°F)]	Design Pressure [MN/m ² (psig)]	Material of Construction or Pump Type	Notes
				S/T	S/T	S/T	
L-701 A/I	First-stage shift converter (HTS)	9	3.35 m ϕ x 16.8 m (11 ft 0 in. ϕ x 55 ft)	694 (790)	4.07 (590)	SA-533 Liner: 316 SS	HTS catalyst bed = 13.7 m (45 ft)
L-702 A/I	Second-stage shift converter (LTS)	9	4.88 m ϕ x 19.2 m (16 ft 0 in. ϕ x 63 ft)	547 (525)	4.07 (590)	SA-299 Liner: 316 SS	LTS catalyst bed = 16.2 m (53 ft)
L-703 A/F	Methanator	6	1.58 m ϕ x 9.14 m (5 ft 2 in. ϕ x 30 ft)	630 (675)	11.13 (1615)	SA-299	Ni catalyst bed = 6.1 m (20 ft)
A-701 A/F	CO ₂ absorber	6	4.57 m ϕ x 27.4 m (15 ft 0 in. ϕ x 90 ft)	330 (135)	10.9 (1580)	SA-299	41 sieve trays
A-702 A/F	CO ₂ stripper	6	4.57 m ϕ x 22.9 m (15 ft 0 in. ϕ x 75 ft)	330 (135)	0.276 (40)	SA-299	27 sieve trays
A-703	Process gas dryer	6	4.27 m ϕ x 6.1 m (14 ft 0 in. ϕ x 20 ft)	330 (135)	4.0 (580)	SA-299	5 sieve trays
A-704	Solvent dryer	6	Top: 4.57 m ϕ x 6.1 m (15 ft 0 in. ϕ x 20 ft) Bottom: 2.18 m ϕ x 3.66 m (7 ft 0 in. ϕ x 12 ft)	477 (400)	0.276 (40)	SA-299	Top: 5 bubble cup trays Bottom: 2 bubble cup trays
T-701	CO ₂ stripper reboiler		--	--	--	--	None in case 5
T-702A A/F	BFW heater	6	Total = 5570 m ² (60,000 ft ²)	547/491 (525/425)	4.03/5.62 (585/815)	2-1/4 Cr - 1 Mo/ 2-1/4 Cr - 1 Mo	2 shells each
T-702B A/F	BFW heater	6	Total = 3440 m ² (37,000 ft ²)	464/450 (375/350)	4.03/5.62 (585/815)	316 SS/ 316 SS	
T-707	CO ₂ stripper condenser		--	--	--	--	None in case 5
P-701	Solvent pump (lean solution or high-pressure solvent)	30	Total = 5.6 m ³ /sec (88,700 gpm)	330 (135)	$\Delta P = 5.13$ (745)	Centrifugal	34 MW (45,600 hp)
P-702	Low-pressure solvent pump	30	Total = 5.6 m ³ /sec (88,700 gpm)	330 (135)	$\Delta P = 5.41$ (785)	Centrifugal	36 MW (48,300 hp)
R-701	Hydrogen compressor (process gas compressor)	1	195 MW (261,000 hp)	334 (250)	$\Delta P = 6.99$ (1015)	Centrifugal	42.0 kg-mole/sec (2,100,000 scfm)
R-702	Hydrogen booster compressor	1	19 MW (25,400 hp)	332 (210)	$\Delta P = 3.55$ (515)	Centrifugal	14.3 kg-mole/sec (715,000 scfm)